1997 December 18

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ALCOA TECHNICAL CENTER
100 TECHNICAL DRIVE • ALCOA CENTER, PA 15069-0001

Development of a High Energy Efficient Pressure Calciner

Final Report for Period
1994 June 01 to 1997 July 31

J. Finley Bush

Work Performed Under
Contract No. DE-FC07-94ID13237

Prepared for
U.S. Department of Energy

Prepared by
Aluminum Company of America
Alcoa Technical Center
100 Technical Drive
Alcoa Center, PA 15069-0001

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ABSTRACT

During the life of this contract, the design, procurement, and construction of a pilot, self-fluidizing, pressure calciner for the production of smelting grade alumina was completed. Initial operating characteristics were determined, and the first half of the first DOX was completed. A design capacity of at least 100 kg/hr of product had been chosen to insure a 100:1 maximum scale-up ratio for the semi-commercial unit. Detailed numerical analysis was made for the heat exchanger design to set the active tube length at 8.5 m (28 ft). The instrumentation and data logging system was designed to obtain the detailed engineering parameters for design of the semi-commercial unit. The pressure feed, discharge, and burner systems were chosen from existing commercial designs to reduce the development work required. Auxiliary equipment, steam condenser, cooling tower, and product cooler, were chosen to simplify operation during the experimental program.

Self-fluidizing capabilities were determined to exist both from temperature profiles and heat transfer coefficient calculations. Operating characteristics of the pilot pressure calciner for the production of smelting grade alumina showed the following:

- The GEMCO valves used in the hot lock hopper discharge system were initially redesigned in an attempt to hold pressure with little or no solids in the chamber and were eventually replaced with a design by Everlasting Valve Company which worked.
- The pulse combustor was operated at both atmospheric and typical operating pressure (7.3 bar) and was the least troublesome portion of the unit.
- Considerable operating problems were found in the set up of the Macawber (hot hydrate feed system) and the controls had to be reprogrammed for dependable operation.
- Condensate was found in the Macawber when hydrate was held at pressure and 180°C for periods greater then 24 hr (bench work established decomposition of the Gibbsite starts at temperatures greater than 140°C when at pressure).
- The ControlView control program by Allen Bradley provided significant flexibility for this project.
- The cyclone dust separator had a limited operating range for efficient solids separation and was replaced with a dust separator with a porous metal filter element and lock hopper.

A limited portion of the DOX was completed and calcination at 150 kg/hr feed (100 kg/hr final product) showed the ability to achieve mono-hydrate (boemite) with an internal tube temperature of greater then 400°C.

INTRODUCTION

The purpose of this report is to cover the work completed during the life of this project for the DOE Metals Initiative Project "Energy Efficient Pressure Calciner." The objective of this project was to take Alcoa's present-day technology and continue to develop it into a viable, commercial process; thus improving the U.S. competitive edge in aluminum production and reducing energy requirements. Since alternate fuels such as coal or oil can be used in this indirect process, a reduction of U.S. dependence on natural gas resources can be realized, if capital and fuel prices warrant the conversion.

The overall goals of this project were to develop sufficient operational and design data (materials of construction, mechanical, etc.) to complete a definitive evaluation of the proposed calciner unit as compared to state-of-the-art (fluid-flash calciner) technology. In addition, project outcomes were to determine the appropriateness for continuing on to Phase IA (Economic Evaluation) and Phase II (Construction and operation of a 10 ton/hr semi-commercial unit).

To meet these goals, successful completion of Phase I was to determine the following:

1) Material can be moved continuously into and out of a pressure vessel with reliability;
2) The SGA produced in a larger unit is equal to or superior to the SGA produced today in the fluid flash calciner, and at least duplicates the material produced in the first single-tube pilot research;
3) Self-fluidization of the materials is achievable on the tube side (Achieved);
4) A mechanical design of the calciner heat exchanger is achievable with either pressure or atmospheric combustion on the shell side (Achieved);
5) Deaeration of the feed is feasible and minimizes the non-condensables in the steam;
6) Define the appropriate construction materials for the tubes, which will minimize corrosion, erosion, and cost; exhibit good heat transfer; and have acceptable life at 850°C shell temperatures (Achieved); and
7) With either pressure or atmospheric combustion, heat transfer coefficients can be achieved to calcine the material at 650°C, while holding internal shell side temperatures to a maximum of 850°C (operation proved unstable above 500°C for 5.08 cm (2 in. diameter) tubes).

SUMMARY OF WORK

Starting with a conceptual design of the commercial facility as shown in Figure 1, the design of the pilot plant, capable of obtaining the information requested in the Statement of Work, Appendix A, was completed. In a commercial unit, wet filter cake is dried in a fluid dryer at 180°C. This removes the moisture and preheats the material to the condensation temperature of the steam leaving the reactor vessel.

The hot, dry material is pressurized and transported with steam into the calciner. The material is calcined by hot flue gas on the outside of the tubes and exits the tubes into a holding vessel to finish calcination. The material is moved through the lock hopper and into a cooler and then into storage. The hot flue gas exits the heat exchanger and part of it is recycled to keep the shell side temperature below the design limit. The remaining flue gas passes through a steam boiler to generate pressurization and transport steam. Steam produced by the calciner is transported to the digestion area and replaces existing steam.

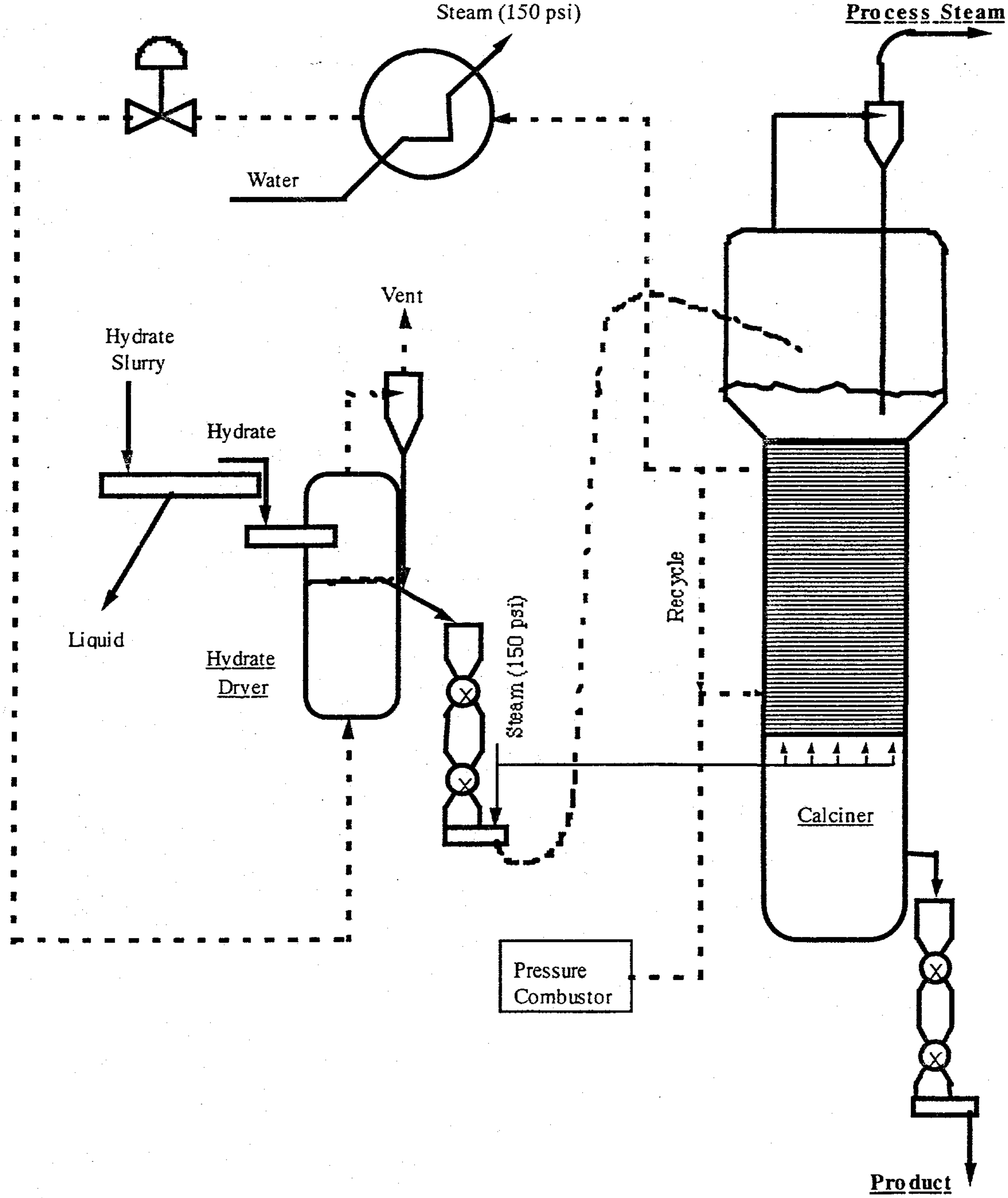


Figure 1: Commercial Design

PROCESS AND INSTRUMENTATION DIAGRAM FOR PILOT PLANT FACILITY

The process flow sheet for the pilot unit is shown in Figure 2. Dry Pt. Comfort hydrate <1> is preheated in a screw heater and charged to a feed tank <2>. The material is pressurized with steam and fed by the Control-veyor through a 12.7 mm line <3> into the pressure calciner. As the material passes down through the tubes, it is fluidized by the chemically bound water as it is released by the heat transferred from the hot shell side. After leaving the tubes the material is retained in a holding vessel <4> at the final exit temperature to finish calcination.

A GEMCO designed valve system <5> is used to drop the pressure and release the solids into a multi-disk cooler to cool the solids. The cooled product <6> is collected in steel drums for sampling and storage.

The steam leaves the calciner <7>, passes through a cyclone and small dust filter to remove particulate, and then is piped to <8> a steam condenser. The condensed steam <10> is separated from the non-condensables <9> and sent to a holding tank. The condenser is cooled by circulating water <19> through a commercial cooling tower.

Commercial natural gas <11> is compressed, mixed with compressed air <13> and burned in the pulse combustor. Skin temperature of the burner is cooled with additional compressed air <14>. The burner recycles cooled combustion gasses <17> by eduction to help control the temperature of the gas <15> to the shell side of the tubes of the calciner. Excess products of combustion (POCs) are vented through a control valve to control the backpressure.

The auxiliary equipment was sized to handle both a three tube 5.08 cm (2 in.) design configuration (installed) and a single 10.2 cm (4 in.) design configuration. The second configuration represents a single tube from the commercial design. This will aid in the design of the commercial unit and facilitate the interpretation of the data for scale-up. The list of equipment is shown in Appendix B - Table 1.

Photographs of the installed equipment are presented in Pictures 1 through 7, Appendix D. The captions are self-explanatory and no additional information is given. The photographs are cross referenced in the list of equipment, Appendix B - Table 1.

To determine the materials of construction to use for the tubes, three different stainless steel tube materials (low, medium, and high cost) were chosen. The choice was based upon the initial cost and resistance to corrosion and erosion. The three tubes used are listed below.

Tube Material	Initial Cost/Meter	Percent Cr
Type 304	$20.11	18
Type 310	26.25	24
Type 347	34.91	17

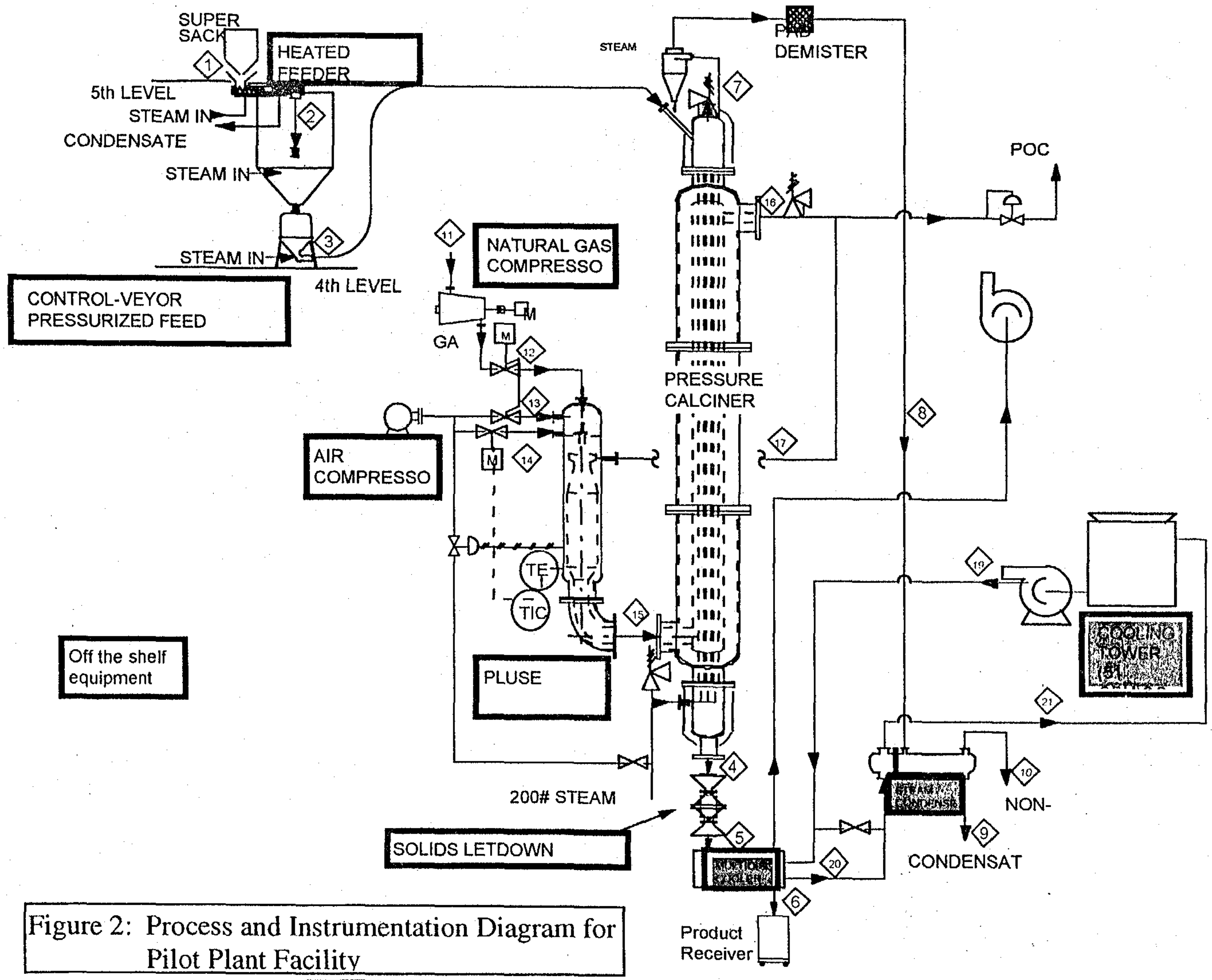


Figure 2: Process and Instrumentation Diagram for Pilot Plant Facility

Everlasting valves were being fabricated. This arrangement was used to do four parts of the designed experiment.

The first mechanical fix was to redesign the seal such that it floated under a pre-set load on the dome valve. This seal, while not perfect, did allow us to operate with tolerable leakage. The second fix was to install a third valve in a second chamber below the GEMCOs.

We were still not satisfied that the operational problems have been solved with the GEMCOs. The valves were sensitive to installation procedures and have a tendency to stick if not used on a day-to-day basis. Another vendor, Everlasting Valve Company, who has helped develop hot discharge solids valves for pressure operation, has been found and that company's valves were installed in early 1997 and operated satisfactorily.

Calciner Unit Design

The castable lined shell for the calciner held up well throughout the life of the project. No hot spots were ever noted on the shell even though the unit was cycled in routine day-to-day operation for the past two years. No tube failures were noted either. During the valve change-out, the expansion joints were inspected and showed no visible signs of cracking.

Although not shown here, due to the crude method of calculation and limited data, the heat transfer coefficients determined are up to 25% lower then in the original design. Thus, it is expected that the unit would need to be larger for the same throughput.

Control System

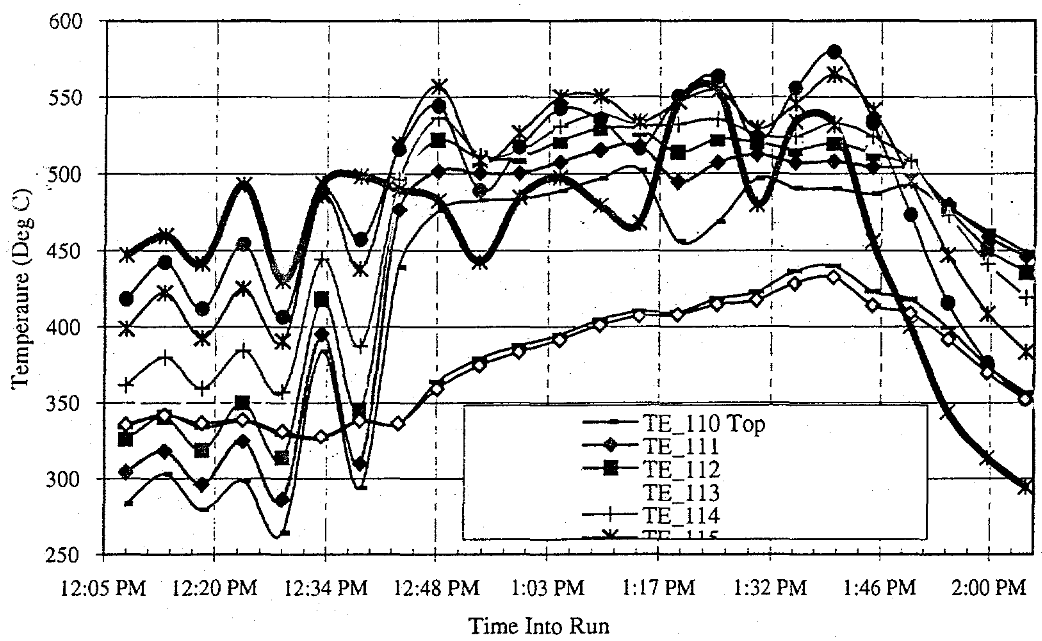
The choice of the Allen-Bradley system using ControlView as the MMI (Man/Machine Interface) has proven itself to be an extremely valuable tool. This system is very versatile and allowed us to make the needed changes as we learned what the true operating requirements were. This versatility however required that we had a trained programmer available on a daily basis to make the necessary changes to the code. The screens used by the operator are shown in Appendix E.

All of the data during a run were stored and available for use in an EXCEL spreadsheet. Macros were written to take the data at the end of a day's run to do the heat and material balances while at the same time testing the raw data for consistency. We anticipated being able to review the day's calculated results to guide us in the next day's operation during the DOX.

Self Fluidization

Finally, we were able to show that self-fluidization was taking place in the tubes by observing temperature profiles in the unit and by the increase in heat transfer coefficient in the tube. Six runs of a nine run DOX were completed before the project was terminated. During this run time it was established that we were not able to maintain stable operation at temperatures above 500°C (see graph below). TE110 through TE117 are the inside tube temperatures. The instability of the temperatures forced an emergency shut down due to excessive tube expansion.

550 Deg C Run



CONCLUSIONS

The unit was near full operational capability prior to the decision to cease work. Although several experiments were completed it is impossible to generate a valid conclusion at this time. However, the limited data did indicate that a run at pressure on both sides of the tubes, self-fluidization, and profiles similar to that initially calculated were achieved. The design and operating teams believe that the project was close to achieving limited success. No further work is contemplated on this project. Alcoa Business Unit management declined to continue funding this project entirely with Alcoa funds for the following reasons:

1. The limited data indicated lower than anticipated heat transfer coefficients and the size of the unit would have to be bigger than originally planned thus impacting economics.
2. The recurrent problems with the feed system.
3. The high cost to build prototype, thus even higher technical and economic risk.

the calciner will be estimated. Material entrainment and its effect on usability of the steam will be assessed.

Subtask 1.2 - Analytical Studies - The participant will reevaluate self-fluidization and solids flow for top feed and bottom discharge to determine tube size and tube length (the original estimate of tube length was 30 ft). Tube sizing will be evaluated based on the fluidization required, heat transfer coefficients, and use of standardized materials to meet code requirements. The effect of tube size on scale-up (number of tubes required) to the 10,000 kg/hr system will also be assessed. Analysis of residence time as a function of tube size, and steam velocities (either induced or self-generated) will be estimated. The participant will perform heat transfer analysis of a three-tube (nominal three-tube) triangulated pitch configuration using a baffled, staggered tube arrangement. The temperature difference as a function of the position will be estimated. Refractory thicknesses necessary to hold the shell temperature to 650°C will be estimated. Chemical evaluation of the changes in converting Gibbsite to SGA will be undertaken as it relates to pressures and temperatures within the fluidized column. As a minimum the participant will define these design review parameters: 1) material temperatures at tube bottom and top, tube diameter, tube thickness, tube length, tube spacing within the shell, shell size and thickness, 2) particle size, particle velocity, particle residence time, 3) steam velocity, and void fraction, and 4) energy requirements (per ton) of product. These results shall be provided to DOE, along with the criteria used to evaluate scale-up from the 100 kg/hr pilot pressure calcination system to the 10,000 kg system will be provided.

Subtask 1.3 - Materials of Construction - Materials of construction will be determined for the calciner tubes. The materials will be evaluated relative to corrosion, erosion, and heat transfer. Materials selection will also be defined for feeders, dischargers, and other ancillary equipment.

Subtask 1.4 - Pilot Plant Component Design - Components for the pilot scale test system will be designed by the participant based on the results of the analytical studies. These components will include multiple tubes for fluidization, a refractory lined outer shell, atmospheric or pressure combustion system, feed and discharge components (solids letdown subsystems), etc. The feed and discharge systems shall be sized large enough to follow the test plan requirements.

Fabrication drawings for the components will be developed for issuance to fabrication vendors. Additionally, specifications for commercially available ancillary components, such as solid feeders and discharges, will be prepared. Specifications for construction will be made. Component operational specifications will be determined and will be utilized during the shakedown tests for confirmation. Pilot plant layout diagrams and piping and instrumentation drawings (P&IDs) will be prepared. Detailed operating instructions for each component will be developed.

Subtask 1.5 - Instrumentation - The design will include sufficient instrumentation to document the parameters of concern. Tube-side and shell wall-side thermocouples; flow monitoring devices for solids and gases; and pressure sensing instrumentation to maintain proper pressure drop throughout the process. All instrumentation will feed information into an automated data collection system. Visual ports, unless excluded from use because of safety constraints, will be

Subtask 3.2 - Data Analysis - The participant will evaluate the data collected during the testing phase. This analysis will define both the production energy requirements and the quality of the alumina product.

Task 4 - Economic and Mechanical Evaluation - The participant will prepare a definitive economic and mechanical evaluation of the semi-commercial (10,000 kg/hr) calciner design and the full-scale-sized unit (50,000 kg/hr). A comparison to start-of-the-art flash calciners will be made. Capital and operating costs will be defined. The test results (energy requirements and product quality obtained during the test) will be used. Where applicable, projected performance (through improvements in design, etc.) will also be used. A risk assessment will be completed of the technology's probability to meet or exceed expectations for a new calciner. Results of the analysis will be used to perform a payback analysis and to assess the associated risks of the technology. The assessment will include evaluation of cost benefits due to dust reduction and alumina quality improvements.

Task 5 - Program Management / Documentation and Reporting - The objective of the Program Management and Reporting task is to provide the overall direction and control necessary to ensure that the proposed program is conducted in a highly professional and cost-effective manner, and to ensure that the program is completed on schedule and within the allocated cost. The participant will conduct technical reviews, coordinate check-point evaluations, and coordinate the technical liaison between DOE and Alcoa.

Subtask 5.1 - Reporting and Documentation - A final report will be prepared that documents the activities in each task and provides project recommendations. A review meeting will be held to determine project status and the advisability of project continuation. A review paper will be presented at an appropriate technical conference or meeting.

APPENDIX B

TABLE I

MAJOR EQUIPMENT LIST

Item No.	Description	Vendor	Capacity	Picture No.
1	Screw Conveyor Heater	WRC Sales	300 kg/hr	3
2	Hydrate Feed System	Macawber	300 kg/hr	3
3	Pressure Calciner	PA Tank & Tube	300 kg/hr	5 & 6
4	Refractory Lining	Chiz Bros.		
5	Fluidizing Tubes	Robert James Co.	See Table pg 9	
6	Tube & Baffle Assembly	CIC		
7	Solids Letdown System	GEMCO	265 kg/hr	6
7	Expansion Joints	Pathway		
8	Multi-Disk Cooler	Heyl & Patterson	265 kg/hr	7
9	Steam Cyclone	Fisher Klosterman	92 kg/hr	2
10	Mist Eliminator	ACS Industries		2
11	Steam Condenser	Hoffman Process	182 kg/hr	
12	Cooling Tower	Diversified Air Systems	195 lpm	
13	Combustion System	MTCI	359,000 kJ/hr	4
14	Gas Compressor	Kruman Equipment	393 slm @ 9.6 bar	
15	Borescope	Olympus America		

TABLE II
Instrumentation List

		Design Values									
Description	Device Type	Flow (kg/hr)	Temp (ºC)	Pressure (kg.cm2)	Sp. Gr.	DP (kg.cm2)	Range	Units	Filud	Function(s)	Device Type
Fluidizing Steam Control Valve	Flow Control Valve	2.3	190.55	10.55		1.06	0 - 5	kg/hr	Steam	Control	Globe Valve
Fluidizing Steam Flow	Flow Element	2.3	190.55	10.55			0 - 5	kg/hr	Steam	Monitor, Record, Totalize	Orifice Plate
Fluidizing Steam Flow	Diff. Press. Transmitter	2.3	190.55	10.55			0 - 5	kg/hr	Steam	Monitor, Record, Totalize	Diff. Pressure Xmtr
Startup Fluidizing Air Pressure Regulator	Pressure Control Valve								Air		Differential Pressure Reducing Regulator
Fluidizing Steam/Startup Air Flow Valve	Solenoid Valve	2.3	204.4	10.55		4.57	0 - 5	kg/cm2	Steam/ Air	Control	Ball Valve
Calciner Pressure Control Valve	Pressure Control Valve	1,861.7	593.33	7.17		8.44	0 - 10	kg/cm2	POC	Control	Butterfly Valve
Calciner Inlet Gas Pressure	Pressure Transmitter		848.88	7.38			0 - 10	kg/cm2	POC	Control, Monitor, Record	Gauge Pressure Xmtr
Calciner Outlet Gas Pressure	Pressure Transmitter		593.33	7.17			0 - 10	kg/cm2	POC	Monitor, Record	Gauge Pressure Xmtr
Calciner Outlet Gas Temperature	Thermocouple	1,861.7	593.33	7.17			-270 - 1370	ºC	POC	Monitor, Record	Thermocouple
Calciner Outlet Gas Temperature	Thermowell										Thermowell
Waste Gas Flow to Stack	Flow Element	544.2	593.3	7.03			0 - 600	kg/hr	POC	Monitor, Record, Totalize	Orifice Plate
Waste Gas Flow to Stack	Diff. Press. Transmitter	544.2	593.3	7.03			0 - 600	kg/hr	POC	Monitor, Record, Totalize	Diff. Pressure Xmtr
Fuel Supply Pressure	Pressure Transmitter		20.0	8.80			0 - 10	kg/cm2	Nat Gas	Alarm, Monitor, Record	Gauge Pressure Xmtr
Waste Gas Stack Temperature	Thermocouple	455.8	593.3	ATM			-270 - 1370	ºC	POC	Monitor, Record.	Thermocouple
Waste Gas Stack Temperature	Thermowell										Thermowell
Combustion Air Pressure	Pressure Transmitter		50.0	8.80			0 - 10	kg/cm2	Air	Alarm, Monitor, Record	Gauge Pressure Xmtr
Fuel Supply/Comb. Chamber Differential Pressure	Diff. Press. Transmitter		848.9	8.80		1.407 (20 lbs)	0 - 10	kg/cm2	POC	Alarm, Control, Monitor, Record	Diff. Pressure Xmtr
Recycle Gas Flow	Flow Element	272.1	593.3	7.03			0 - 300	kg/hr	POC	Control, Monitor, Record	Annubar
Recycle Gas Flow	Diff. Press. Transmitter	272.1	593.3	7.03			0 - 300	kg/hr	POC	Control, Monitor, Record	Diff. Pressure Xmtr

TABLE II (cont.)

Description	Device Type	Design Values					Range	Units	Fluid	Function(s)	Device Type
		Flow (kg/hr)	Temp (°C)	Pressure (kg.cm2)	Sp. Gr.	DP (kg.cm2)					
Heated Screw Feeder Steam Inlet Valve	Solenoid Valve	45.4	388.0	14.07		1.76	0 - 20	kg/cm2	Steam	Control	Solenoid Valve
Roto-Feed Discharge Pressure	Pressure Indicator		160.0	11.25			0 - 15	kg/cm2	Alumina		Pressure Gauge
Heated Screw Feeder Discharge Temperature	Thermocouple	264.4	148.9	ATM			-270 - 1370	ºC	Al Hyd.	Monitor, Record	Thermocouple
Heated Screw Feeder Discharge Temperature	Thermowell										Thermowell
Feed System Discharge Temperature	Thermocouple	355.1	160.0	14.07			-270 - 1370	ºC	Al Hyd.	Control, Monitor, Record	Thermocouple
Feed System Discharge Temperature	Thermowell										
Calciner Hydrate Inlet Feed Temperature	Thermocouple	355.1	160.0	14.07			-270 - 1370	ºC	Al Hyd.	Alarm, Monitor, Record	Thermocouple
Calciner Hydrate Inlet Feed Temperature	Thermowell										
Conveying Steam Flow	Flow Element	90.7	221.1	14.07	N/A	N/A	N/A	N/A	Steam	Alarm, Monitor, Record, Totalize	Orifice Plate
Conveying Steam Flow	Diff. Press. Transmitter	90.7	221.1	14.07	N/A	N/A	N/A	N/A	Steam	Alarm, Monitor, Record, Totalize	Diff. Pressure Xmtr
Conveying Steam Pressure Regulator	Pressure Control Valve	90.7	221.1	14.07		3.52	0 - 20	kg/cm2	Steam		Pressure Reducing Regulator
Conveying Line Steam Pressure Regulator	Pressure Control Valve		388.0	14.07			0 - 20	kg/cm2	Steam		Pressure Reducing Regulator
Calciner Lower Bonnet Material High Level	Level Switch		648.9	9.49				%	Alumina	Alarm, Monitor, Record	Vibration Level Wand
Calciner Lower Bonnet Material Low Level	Level Switch		648.9	9.49				%	Alumina	Alarm, Monitor, Record	Vibration Level Wand
Solids Letdown System Pressure High	Dual Pressure Switch		648.9	9.49			0 - 15	kg/cm2	POC	Control, Monitor, Record	Pressure Switch
Solids Letdown System Pressure Low	Dual Pressure Switch		648.9	9.49			0 - 15	kg/cm2	POC	Control, Monitor, Record	Pressure Switch
Product Discharge Pressure	Pressure Transmitter		648.9	9.49			0 - 15	kg/cm2	Alumina	Monitor, Record	Gauge Pressure Xmtr
Product Discharge Temperature @ Calciner	Thermocouple	185.9	648.9	9.49			-270 - 1370	ºC	Alumina	Monitor, Record	Thermocouple
Product Discharge Temperature @ Calciner	Thermowell										Thermowell

TABLE II (cont.)

Description	Device Type	Design Values: Flow (kg/hr)	Temp (ºC)	Pressure (kg.cm2)	Sp. Gr.	DP (kg.cm2)	Range	Units	Fluid	Function(s)	Device Type
Solids Letdown System Inlet Gate Control	Solenoid Valve	N/A	N/A	5.63 - 7.03 (80-100 PSIG)		N/A	5.63 - 7.03 (80-100 PSIG)	kg/cm2	Instr Air	Control	Solenoid Valve
Solids Letdown System Outlet Gate Control	Solenoid Valve	N/A	N/A	5.63 - 7.03 (80-100 PSIG)		N/A	5.63 - 7.03 (80-100 PSIG)	kg/cm2	Instr Air	Control	Solenoid Valve
Solids Letdown System Pressurize Valve	Solenoid Valve	185.9	648.9	9.49		9.49	0 - 15	kg/cm2	POC	Control	High Temp Ball Valve
Multidisc Cooler Inlet Chute Full	Level Switch		648.9	ATM				%	Alumina	Alarm, Monitor, Record	Vibration Level Wand
Solids Letdown System Vent Valve	Solenoid Valve	185.9	648.9	9.49		9.49	0 - 15	kg/cm2	POC	Control	High Temp Ball Valve
Multidisc Cooler Temperature #1	Thermocouple	185.9	648.9	ATM			-270 - 1370	ºC	Alumina	Monitor, Record	Thermocouple
Multidisc Cooler Temperature #2	Thermocouple	185.9	648.9	ATM			-270 - 1370	ºC	Alumina	Monitor, Record	Thermocouple
Multidisc Cooler Temperature #3	Thermocouple	185.9	648.9	ATM			-270 - 1370	ºC	Alumina	Monitor, Record	Thermocouple
Multidisc Cooler Flow Indicator #1	Rotameter	3785 L/hr	29.4	2.83			0 - 4000	L/hr	Water	Indicate	Variable Area Flowmeter
Multidisc Cooling Water Outlet Temperature Indicator #1	Temp. Indicator w/well		32.2	2.46			0 - 50	ºC	Air		Bi-Metal Thermometer
Multidisc Cooler Flow Indicator #2	Rotameter	3785 L/hr	29.4	2.83			0 - 4000	L/hr	Water	Indicate	Variable Area Flowmeter
Multidisc Cooling Water Outlet Temperature Indicator #2	Temp. Indicator w/well		32.2	2.46			0 - 50	ºC	Air		Bi-Metal Thermometer
Multidisc Cooler Flow Indicator #3	Rotameter	3785 L/hr	29.4	2.83			0 - 4000	L/hr	Water	Indicate	Variable Area Flowmeter
Multidisc Cooling Water Outlet Temperature Indicator #3	Temp. Indicator w/well		32.2	2.46			0 - 50	ºC	Air		Bi-Metal Thermometer
Steam Cyclone High Differential Pressure	Pressure Differential		204.4	8.44			0 - 15	kg/cm2	Steam	Alarm, Monitor, Record	Diff. Pressure Switch
Steam Cyclone Discharge Pressure	Pressure Transmitter		204.4	8.44			0 - 15	kg/cm2	Steam	Alarm, Monitor, Record	Gauge Pressure Xmtr

TABLE II (cont.)

Description	Device Type	Design Values Flow (kg/hr)	Temp (°C)	Pressure (kg.cm2)	Sp. Gr.	DP (kg.cm2)	Range	Units	Fliud	Function(s)	Device Type
Steam Cyclone Discharge Temperature	Thermocouple	172.3	204.4	8.44			-270 - 1370	°C	Steam	Monitor, Record	Thermocouple
Steam Cyclone Discharge Temperature	Thermowell										Thermowell
Non-Condensable Flow	Flow Element	0.17 Nm3/hr	93.3	8.44			0 - 0.5	Nm3/hr	Air	Monitor, Record, Totalize	N/A
Non-Condensable Flow	Diff. Press. Transmitter	0.17 Nm3/hr	93.3	8.44			0 - 0.5	Nm3/hr	Air	Monitor, Record, Totalize	Turbine Flowmeter
Non-Condensate Pressure Control Valve	Pressure Control Valve	172.3	204.4	8.44		8.44			Air	Control	Globe Valve
Condenser Steam Side Pressure	Pressure Transmitter		204.4	8.44			0 - 10	kg/cm2	Steam	Monitor, Record	Gauge Pressure Xmtr
Steam Condenser Non Condensables Temperature	Thermocouple	172.3	204.4	8.44			-270 - 1370	°C	Air	Monitor, Record	Thermocouple
Steam Condenser Non Condensables Temperature	Thermowell										Thermowell
Condensate Flow From Condenser Volume Tank	Flow Element	170.325 L/hr	93.3	8.44			0 - 200	L/hr	Water	Monitor, Record, Totalize	Turbine Flowmeter
Condensate Flow From Condenser Volume Tank	Diff. Press. Transmitter	170.3	93.3	8.44			0 - 200	kg/hr	Water	Monitor, Record, Totalize	N/A
Condensate Volume Tank Level Control Valve	Level Control Valve	170.3	93.3	ATM		8.44			Water	Control	Globe Valve
Condensate Volume Tank Level	Level Transmitter	170.3	93.3	ATM				%	Water	Monitor, Record	Capacitance
Condensate Volume Tank Level	Diff. Press. Transmitter	170.3	93.3	ATM				%	Water	Monitor, Record	N/A
Demister High Differential Pressure	Pressure Differential		204.4	8.44			0 - 15	kg/cm2	Steam	Alarm, Monitor, Record	Diff. Pressure Switch
Cooling Tower Inlet Temperature	Thermocouple	11355 L/hr	43.3	1.76			-270 - 1370	°C	Water	Monitor, Record	Thermocouple
Cooling Tower Inlet Temperature	Thermowell										Thermowell
Cooling Water Flow	Rotameter	11355 L/hr	29.4	2.83			0 - 15000	L/hr	Water	Indicate	Variable Area Flowmeter
Cooling Water Pump #1 Pressure	Pressure Indicator		29.4	2.83			0 - 5	kg/cm2	Water		Pressure Gauge

TABLE II (cont.)

		Design Values									
Description	Device Type	Flow (kg/hr)	Temp (ºC)	Pressure (kg.cm2)	Sp. Gr.	DP (kg.cm2)	Range	Units	Fllud	Function(s)	Device Type
Cooling Tower Outlet Temperature	Thermocouple	3785 L/hr	29.4	2.83			-270 - 1370	ºC	Water	Monitor, Record	Thermocouple
Cooling Tower Outlet Temperature	Thermowell										Thermowell
Cooling Water Pump #2 Pressure	Pressure Indicator		29.4	2.83			0 - 5	kg/cm2	Water		Pressure Gauge
Multidisc Cooler Outlet Cooling Water Temperature	Thermocouple	3785 L/hr	32.2	2.46			-270 - 1370	ºC	Water	Alarm, Monitor, Record	Thermocouple
Multidisc Cooler Outlet Cooling Water Temperature	Thermowell										Thermowell

APPENDIX C

Thermal and Mechanical Design Calculations for a Self-Fluidizing Pressurized Combustion Calciner

Prepared for

Aluminum Company of America

Prepared by

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Pittsburgh, Pennsylvania

Rev. 1 - September 30, 1994

Summary

This report summarizes the assumptions and analyses performed to arrive at the final design of the pressurized combustion calciner pilot plant. This final report has two purposes:

- To clearly delineate the expected performance based on analysis. The analytical methods used are approximate. We want to understand how well they apply to the pilot calciner unit as an aid to ultimate scale-up to the commercial calciner.

- To provide a guide for developing start-up procedures. It is possible that the calciner tubes can be overheated in certain operating upset conditions. This could result in damage to the bellows expansion joints. The tube temperatures are monitored by thermocouples. We need to develop an operating envelope of safe tube temperature profiles. If the envelope is closely approached we can begin to shut the calciner down before any damage is done. It is important to recognize that the calciner vessel pressure boundary is designed to accommodate a worst case failure of the expansion joints. There can never be a safety problem caused by expansion joint failure.

Table of Contents

1.0 Introduction

Basic Technology, Inc. has been commissioned by Alcoa to provide thermal, mechanical, and process design services for the construction of a pressurized combustion alumina trihydrate calculation pilot plant. This report presents the thermal calculations performed as a part of the design effort.

1.1 Conclusions

- For 30 ft long calciner tubes an average outside tube film coefficient of 14 Btu/hr-ft^2-°F is required to process the maximum inlet flow rate of 265 kg/hr of alumina trihydrate. For three 2 in. tubes, average cross flow velocity in a baffled overall cross flow heat exchanger is 13 ft/sec. The total flue gas flow rate through the calciner is sufficient to provide the required film coefficient. For a single 4 in. tube, the average cross flow velocity would have to be approximately 34 ft/sec. Performance with the maximum inlet flow rate of alumina trihydrate is questionable.

- Pressure drop due to tube cross flow is small. The major pressure losses in the recycle system will be due to the baffles, connecting piping, and entrance and exit effects. We can get a better analytical estimate, but previous design experience is the best guide at this time. A 3 to 5 psi pressure drop is a reasonable design criterion for the recycle system.

- System heat losses affect the choice of combustor excess air and recycle ratio. Design of the internal insulation system should be completed to be certain that the initial approximations are satisfactory.

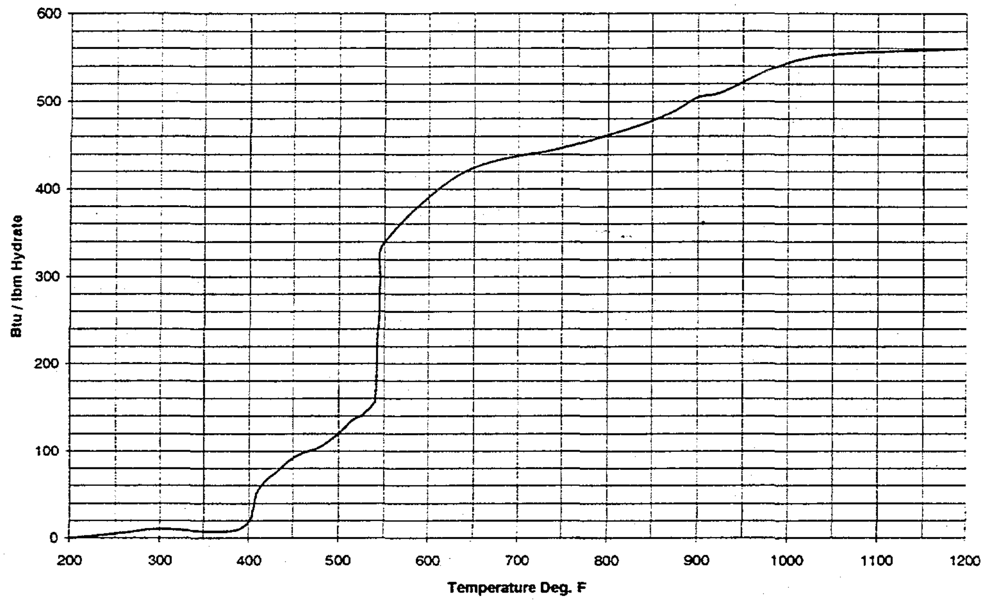
2.0 Process Description

Dehydration of alumina trihydrate is performed in a self fluidized tubular reactor. The dehydration reaction is:

1) $Al_2O_3 \cdot 3H_2O \rightarrow Al_2O_3 + 3H_2O(g)$

The reaction is endothermic and temperatures remain below the $\gamma \rightarrow \alpha$ transition temperature. Process heat requirements are based on cumulative heat of dehydration vs. temperature data supplied by Alcoa as shown in Figure 1. A process overview is shown in Figure 2.

Figure 1 : Dehydration Curve



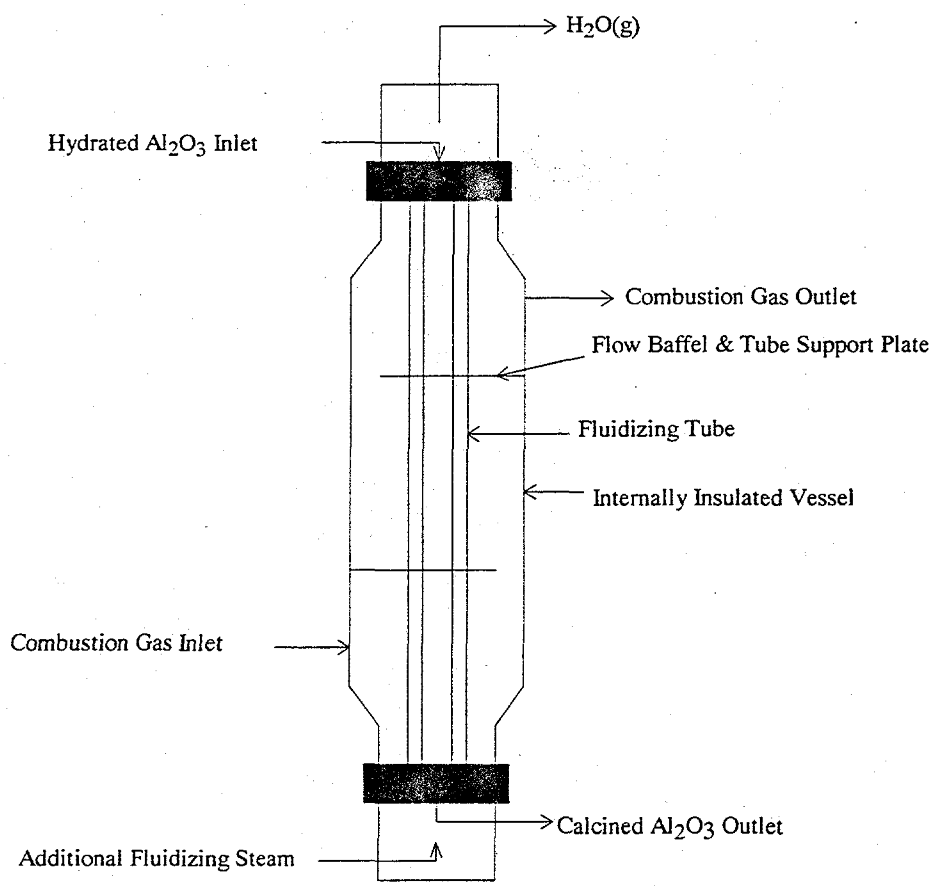


Figure 2: Calciner Vessel

3.0 Reactor Thermal Design

Steam evolved during the dehydration process is vented along with any added fluidizing steam. The molecular weight of the inlet hydrated alumina is:

$$M_w = (2)(27)_{Al} + (6)(16)_O + (6)(1)_H = 156 \frac{Kg}{Kg - Mole}$$

The molecular weight of the evolved steam is:

$$M_w = (2)_H + (16)_O = 18 \frac{Kg}{Kg - Mole}$$

Three moles of steam are produced from each mole of inlet hydrated alumina. Therefore:

$$\frac{(3)(18)}{154} \times 100 = 35 \; percent$$

of the inlet mass leaves as steam. For an inlet flow rate of 265 kg/hr, the steam flow rate is 93 kg/hr (205 lb/hr).

3.1 Preliminary Calculations

The pilot calciner will be operated with three 2 in. diameter tubes and a single 4 in. diameter tube. The initial calculations used tubing dimensions. Pipe material will be used for construction and the calculations are updated to include actual dimensions in this report.

Steam passes through the calciner tubes in a direction counterflow to the inlet alumina. The counterflowing steam must ultimately be accounted for in the heat exchanger design calculations when we address questions of fluidization and tube temperature profiles. However, for the moment we will neglect it. From Figure 1, the total heat input required to heat one pound of the input hydrate from 175°C to 650°C (350°F to 1200°F) is approximately 550 Btu/lb. For the inlet hydrate flow rate of 265 kg/hr we have:

$$\dot{Q}_{in} = (265 \frac{kg}{hr})(2.205 \frac{lbm}{kg})(550 \frac{Btu}{lbm}) = 321379 \frac{Btu}{lbm}$$

This is very close to the contract specified heat rate of 340,000 Btu/hr, and we will use this value for preliminary sizing purposes. Heat exchanger sizing calculations are iterative because many of the phenomena involved are nonlinear. With this caveat, we will make the following initial assumptions:

1. The fluidized particle film coefficient on the inside of the tube is very large and its contribution to the radial thermal resistance can be neglected.

2. The dryer tubes are thin. For a 2 in. external diameter tube, BWG Gage 13, the wall thickness is 0.095 in.. For a 200 psig internal working pressure, the hoop stress is conservatively given using the external radius by:

$$\sigma = \frac{\Pr}{t} = \frac{(200)(1)}{0.095} = 2100\ psi$$

For a two in. pipe the outside diameter is 2.375 in. and the wall thickness is 0.154 in.. This is still quite thin and the increased thickness is offset to some extent by the increased diameter and heat transfer area of the pipe.

For a 4 in. external diameter tube, the stress would only be 4200 psi. Even though we have not selected the precise tube material, it its clear that the hoop stresses will be low and wall thicknesses thin. Therefore, we will also neglect the through wall conductive thermal resistance.

3. The first two assumptions imply that the heat transfer rate will be controlled by the external tube film coefficient.

To obtain an initial estimate of the film coefficient required, we will make additional assumptions:

1. Tube length = 27 ft
2. Combustion gas inlet temperature = 650°C (1560°F)
3. Combustion gas outlet temperature = 540°C (1000°F)
4. Average shell side temperature = 595°C (1280°F)
5. Product inlet temperature = 175°C (350°F)
6. Product outlet temperature = 650°C (1200°F)
7. Average product temperature = 412°C (775°F)

For three 2 in. diameter 30 ft long Schedule 40 pipes, the external surface area is:

$$A = (3)(27)(\pi)(\frac{2.375}{12}) = 50 ft^2$$

For one 4 in. diameter Schedule 40 pipe the surface area is:

$$A = (27)(\pi)\left(\frac{4.5}{12}\right) = 32\ \text{ft}^2$$

The required average unit surface conductance is found from:

$$\dot{Q} = A\bar{h}\Delta T_{avg}$$

$\therefore$

$$\bar{h} = \frac{\dot{Q}}{A\Delta T_{avg}}$$

$\therefore$

For the 2 in. pipes we have:

$$\bar{h} = \left(340{,}000\frac{BTU}{hr}\right)\left(\frac{1}{50\,ft^2}\right)\left(\frac{1}{505^\circ F}\right) \approx 14\frac{BTU}{hr-ft^2-^\circ F}$$

For the 4 in. diameter pipe:

$$\bar{h} = \left(340{,}000\frac{BTU}{hr}\right)\left(\frac{1}{32\,ft^2}\right)\left(\frac{1}{505^\circ F}\right) \approx 21\frac{BTU}{hr-ft^2-^\circ F}$$

The total heat transfer rate and the assumed temperature drop of the combustion gas fixes the required combustion gas flow rate.

$$\dot{Q} = \dot{m}C_p\Delta T$$

Assuming air properties for the combustion gas we have:

$$C_p = 0.27\frac{Btu}{lbm-^\circ F} \text{ at } 1280^\circ\text{F}$$

For $\Delta T_{avg} = 505\ F$

$\therefore$

$$\dot{m} = \frac{\dot{Q}}{C_p\Delta T} = (340{,}000\frac{Btu}{hr})(\frac{hr}{3600\,\text{sec}})(\frac{lbm-^\circ F}{0.27\,Btu})(\frac{1}{560^\circ F}) = 0.625\frac{lbm}{\text{sec}}$$

$$= 2250\frac{lbm}{hr}$$

The volumetric flow rate of combustion gas at standard conditions is:

$$\dot{v} = \frac{\dot{m}RT}{P}$$

where P = 14.7 psia
T = 60°F (520°R)
R = 53.34 ft-lbf/lbm - °R

$$\dot{v} = (0.625\frac{lbm}{\sec})(53.34\frac{ft-lbf}{lbm-^{\circ}R})(520^{\circ}R)(\frac{ft^2}{(14.7)(144)lbf})$$

$$= 8.2\frac{ft^3}{\sec}$$

$$= 492\ SCFM$$

Next we consider the flow velocities required to obtain the calculated unit surface conductances. There are three 2 in. pipes set in a triangular configuration. However, for the moment we will consider a single pipe to obtain some approximate results. The simplified configuration is shown in Figure 3.

For the single tube case, the Nusselt and Reynold's Numbers are related by the correlation:

$$Nu = C\,\mathrm{Re}^n$$

where C = 0.174 and n = 0.618 for $4000 \leq \mathrm{Re} \leq 40000$.

$$Nu = \frac{\bar{h}D_0}{k}$$

where k = 0.036 Btu/hr-ft-°F for air at 1280 °F.

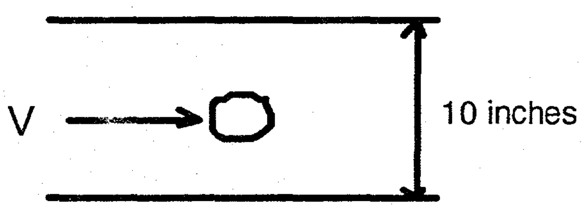


Figure 3: Simplified Tube Geometry

$\therefore$

For 2 in. diameter Schedule 40 pipe:

$$Nu = (14\frac{Btu}{hr-ft^2-°F})(\frac{2.375}{12}ft)(\frac{hr-ft-°F}{0.036Btu})$$
$$= 77$$

The required tube Reynold's Number is:

$$\text{Re} = (\frac{Nu}{C})^{\frac{1}{n}} = (\frac{77}{0.174})^{\frac{1}{0.618}} = 19119$$

In calculating the Reynold's and Nusselt Numbers, we will make pressure corrections for the combustion gas density but will assume the transport properties are dependent only on temperature. For T = 1280°F (1740 °R) and P = 105 psig (120 psia):

$$\rho = \frac{P}{RT} = (120\frac{lbf}{in^2})(144\frac{in^2}{ft^2})(\frac{lbm-°R}{53.34\,ft-lbf})(\frac{1}{1740°R})$$
$$= 0.186\frac{lbm}{ft^3}$$

$$\mu = 2.74\times10^{-5}\frac{lbm}{ft-\sec}$$

$\therefore$

$$V = \frac{\mu\,\text{Re}}{\rho D_0} = (2.74\times10^{-5}\frac{lbm}{ft-\sec})(19119)(\frac{ft^3}{0.186\,lbm})(\frac{12}{2.375\,ft})$$
$$= 14.23\frac{ft}{\sec}$$

This implies a cross flow mass flow rate per ft of exchanger length equal to:

$$\dot{m} = \rho VA = (0.186\frac{lbm}{ft^3})(14.23\frac{ft}{\sec})(0.833\,ft^2)$$
$$= 2.2$$

For a 4 in. diameter pipe we have:

$$Nu = (21\frac{Btu}{hr - ft^2 -{}^\circ F})(\frac{4.5}{12}ft)(\frac{hr - ft -{}^\circ F}{0.036 Btu})$$
$$= 219$$

$$Re = (\frac{Nu}{C})^{\frac{1}{n}}$$

where C = 0.0239 and n = 0.805 for $40{,}000 \le Re \le 400{,}000$.

$\therefore$

$$Re = (\frac{219}{0.0239})^{\frac{1}{0.805}} = 83521$$

$\therefore$

$$V = (2.74 \times 10^{-5}\frac{lbm}{ft - \sec})(83521)(\frac{ft^3}{0.186\ lbm})(\frac{12}{4.5 ft})$$
$$= 33\frac{ft}{\sec}$$

This implies a cross flow mass velocity of:

$$\dot{m} = 5.1\frac{lbm}{\sec}$$

These initial calculations are qualitative because the model used is very simple. However, they are useful when properly interpreted. Table 1 shows the ratio of cross flow mass flow rate to inlet mass flow rate for the 2 in. and 4 in. pipe diameters.

Table 1: Preliminary Ratio of Cross Flow Mass Velocity to Inlet Mass Velocity

2 in. Diameter Pipe Mass Flow Ratio	4 in. Diameter Pipe Mass Flow Ratio
3.5:1	8:1

The mass flow ratios considered across the longitudinal centerline of the vessel do not account for area reductions due to the presence of the tubes. Accounting for the tubes will reduce these ratios. The preliminary conclusion is that the assumed combustion gas flow rate and temperature drop will provide adequate performance with 27 ft long 2 in. diameter pipes, although closely spaced baffles may be required. Performance with the 4 in. diameter pipe is questionable.

3.2 Design Calculations for Three Tube Configuration

The hydrated alumina flows through three 2 in. diameter pipes set on a 5 in. triangular pitch. The pipes are contained in an internally insulated shell with a 10 in. diameter free flow cross section. The general configuration is shown in Figure 4.

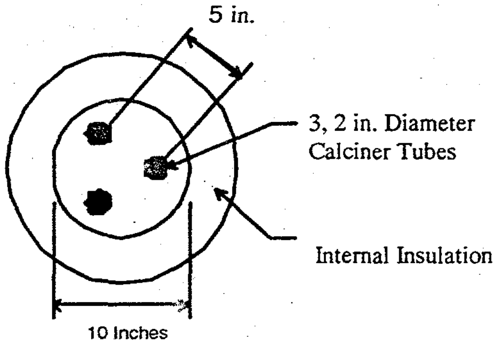


Figure 4: Calciner Cross Section

Preliminary calculations indicate that an average outside tube surface conductance of $14 \dfrac{Btu}{hr - ft^2 - {}^\circ F}$ is required. An initial estimate of the cross flow mass velocity was obtained for a single tube. In this section of the report we consider the effect of the staggered tube arrangement. For staggered tubes, the velocity used in the Reynold's Number is based on the minimum free flow area available for fluid flow. Pertinent dimensions are shown in Figure 5.

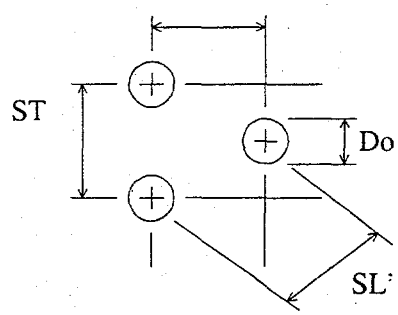


Figure 5: Staggerd Tube Dimensions

where C=0.513
n=0.561, Reference 1, Table 6-4

$\therefore$

$$R_e = \left[\frac{N_u}{CP_r^{\frac{1}{3}}}\right]^{\frac{1}{n}}$$

Where $N_u = \dfrac{hD_o}{k}$

for $h = 14\dfrac{Btu}{hr - ft^2 - {}^\circ F}$

$$N_u = (14\frac{Btu}{hr - ft^2 - {}^\circ F})(\frac{2.375}{12}ft)(\frac{hr - ft - {}^\circ F}{0.036Btu})$$

$$= 77$$

$$R_e = \left[\frac{77}{(0.513)(0.728)^{\frac{1}{3}}}\right]^{\frac{1}{0.561}} = 9149$$

$$R_e = \frac{\rho VD_0}{\mu} = \frac{G_{max}D_o}{\mu}$$

$\therefore$

$$G_{max} = (2.77 \times 10^{-5}\frac{lbm}{ft - sec})(\frac{12}{2.375 ft})(9149)$$

$$= 1.28\frac{lbm}{ft^2 - sec}$$

The total cross flow area is approximately:

$$A = \frac{(3)(2.08)}{12} = 0.52 ft^2 \textit{ per} \text{ foot of exchanger length}$$

Therefore, the total cross flow mass flow rate is:

$$\dot{m} = 0.67\frac{lbm}{sec}$$

These simple considerations would imply a baffle spacing of approximately 1 ft. An actual baffle spacing of 18 in. will likely provide an adequate external film coefficient.

The frictional pressure drop per pass is approximately:

$$\Delta P = \frac{f\, G_{max}^{2} N}{\left(2.09 \times 10^{8}\right)(\rho)} \left(\frac{\mu_s}{\mu_b}\right)^{0.14}$$

$$f' = \left[0.25 + \frac{0.118}{\left(\frac{S_T - D_o}{D_o}\right)^{1.08}}\right] \left(R_{e\max}\right)^{-0.16}$$

$$\mu_s \approx \mu_b = 2.77 \times 10^{-5} \frac{lbm}{ft - \sec}$$

$$D_o = \frac{2.375}{12} = 0.198\, ft.$$

$$S_T = S_L = \frac{4.45}{12} = 0.37\, ft.$$

$$G_{max} = 1.28 \frac{lbm}{ft^2 - \sec} = 4608 \frac{lbm}{hr - ft^2}$$

$$\rho = 0.186 \frac{lbm}{ft^3}$$

$\therefore$

$$f' = \left[0.25 + \frac{0.118}{\left(\frac{0.17}{0..20}\right)^{1.08}}\right] (9149)^{-0.16}$$

$$= 0.091$$

$\therefore$

$$DP = \frac{(0.091)(4608)^2(2)}{(0.186)\left(2.09 \times 10^{8}\right)}$$

$$= 0.093 \frac{lbf}{ft^2}$$

$$DP \; for \; 27 \; passes \; = \; 2.51 \frac{lbf}{ft^2} \approx 0.6 \; inches \; H_2O$$

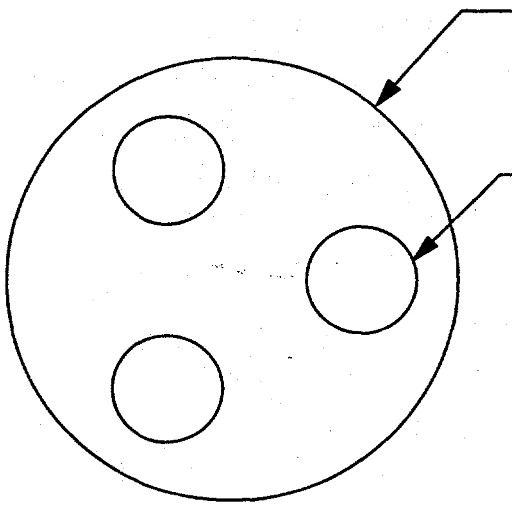


Figure 7: Calciner Cross Section

The pressure drop due to the baffles can be approximated by treating the baffle as an orifice plate in a pipe with a hydraulic diameter equal to that of the calciner axial flow area as shown in Figure 7.

The hydraulic diameter, D_H, of the free flow passage is:

$$
\begin{aligned}
D_H &= 4\frac{\textit{flow area}}{\textit{wetted perimeter}} \\
&= \frac{4\left(\frac{P}{4}\right)\left(D_f{}^2 - D_p{}^2\right)}{P\left(D_f + 3D_t\right)} \\
&= \frac{D_f{}^2 - 3D_p{}^2}{D_f + 3D_t} \\
&= \frac{10^2 - (3)(2.375)^2}{10 + (3)(2.375)} \\
&= 4.85\ in.
\end{aligned}
$$

The flow baffles are clipped as shown in Figure 8.

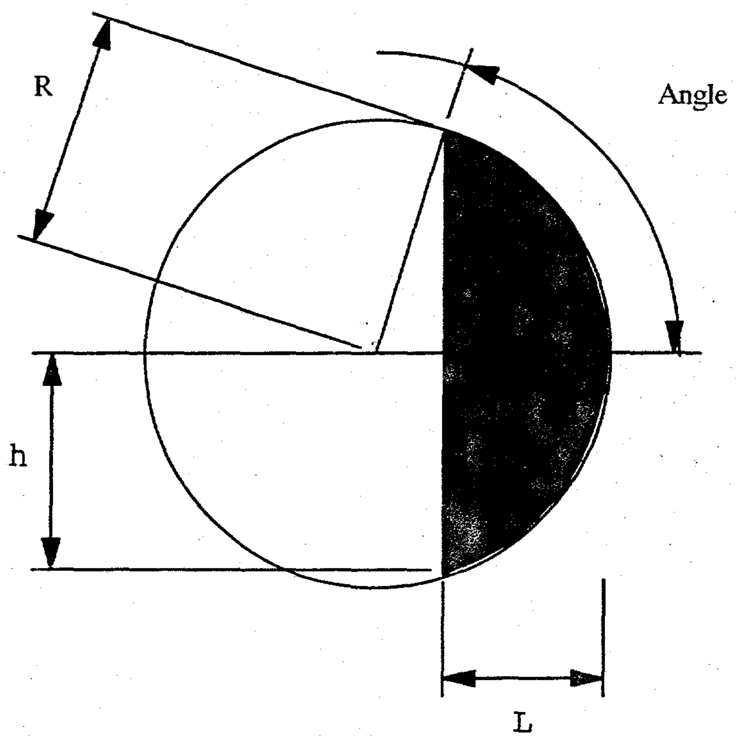


Figure 8: Clipped Baffle

The length, h, is found from:

$$(R - L)^2 + h^2 = R^2$$

$$\therefore$$

$$h = \sqrt{2RL - L^2}$$

The baffles are clipped on alternate sides over 35% of the diameter as shown in Figure 9.

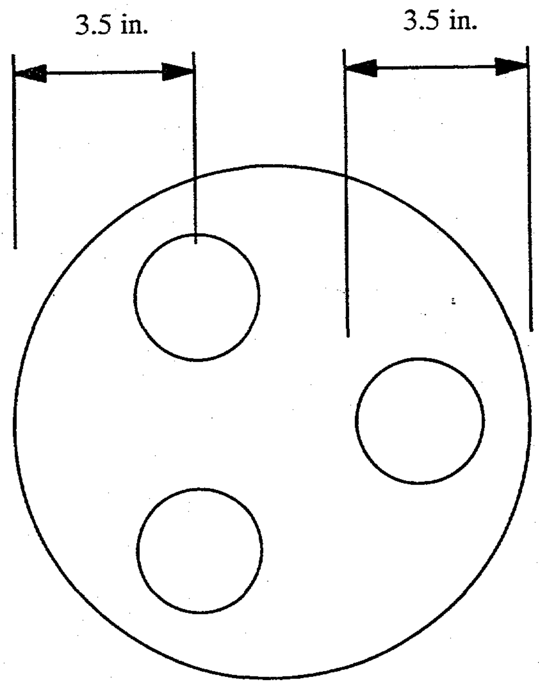


Figure 9: Baffle Clipping Pattern

Therefore:

$$L = 3.5\ in.$$
$$R = 5.0\ in.$$
$$h = 4.77\ in.$$

The clip angle is :

$$q = \cos^{-1}\frac{R-L}{R}$$
$$= \cos^{-1}\frac{1.5}{5} = 72.54$$

The free flow area of the baffle cut is:

$$A_f = \pi R^2\left(\frac{2\theta}{360}\right) - h(R-L)$$
$$= \pi(5)^2\left(\frac{145.08}{360}\right) - (4.77)(5-3.5)$$
$$= 24.49\ in.^2$$

The static pressure drop due to flow across a single baffle is:

$$DP_{baffle} = K\frac{rV^2}{2g_c}$$

where $K = a\left(\frac{D_h}{d_h}\right)\left[1-\left(\frac{d_h}{D_h}\right)^4\right]\left(\frac{1}{C^2}\right)$, Ref.2

where α and C are found from Tables 2 and 3.

Table 2: α vs. Hydraulic Diameter Ratio

d/D	0.2	0.3	0.4	0.5	0.6	0.7	0.8	0.9
α	0.93	0.89	0.82	0.74	0.63	0.53	0.38	0.22

Table 3: Discharge Coefficient, C

	Discharge Coefficient, C							
	Orifice Reynolds Number, $\frac{rVD_h}{m}\cdot\frac{D_h}{d_h}$							
Diameter Ratio d/D	10	60	100	500	10^3	10^4	10^5	10^6
0.3	0.47	0.64	0.67	0.72	0.70	0.60	0.60	0.60
0.5	0.46	0.66	0.69	0.74	0.72	0.61	0.60	0.60
0.7	0.42	0.67	0.72	0.81	0.83	0.65	0.61	0.60

$$\frac{D_h}{d_h} = \frac{4.85}{2.71} = 1.78$$

$\therefore$

$$\alpha = 0.69$$
$$C = 0.60$$
$$K = (0.69)(1.78)^2\left[1-(0.58)^4\right]\left(\frac{1}{0.60}\right)^2$$
$$= 5.40$$

$\therefore$

$$DP_{baffle} = \left(\frac{5.4}{2}\right)\left(7.44\frac{ft}{\sec}\right)^2\left(0.222\frac{lbm}{ft^3}\right)\left(\frac{lbf-\sec^2}{32.174\ lbm\text{-}ft}\right)$$
$$= 1.03\frac{lbf}{ft^2}$$
$$= 7.2\times10^{-3}\ psi\ per\ segment$$

The static pressure drop per segment across the calciner is:

$$DP = 6.5\times10^{-4} + 0.032 + 7.2\times10^{-3}$$
$$\approx 0.04\ psi\ per\ segment$$

We are going to use 17 baffles, so there will be 16 segments. The calculated total pressure drop across the calciner is:

$$DP = 0.64\ psi$$

The pressure drop calculation is admittedly crude, and there will be additional pressure losses in the system piping. However it is important to have an approximate value. The recycle compressor will provide a 3 psi static pressure boost. This simple calculation suggests that the boost will be adequate.

3.3 External Heat Losses

An initial estimate of the process heat losses and calciner vessel temperature was based on the very simple model shown in Figure 10. A more detailed analysis was performed for the design of the vessel internal insulation.

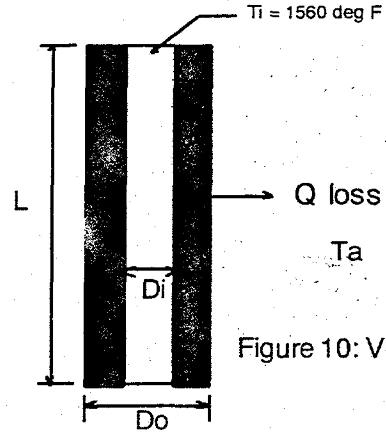


Figure 10: Vessel Heat Losses

The total heat loss is:

$$\dot{Q} = \frac{T_i - T_a}{\frac{1}{h_i A_i} + \frac{\ln(\frac{r_o}{r_i})}{2\pi kL} + \frac{1}{h_o A_0}}$$

The outside surface film coefficient h_o has convective and radiative components,

$$h_0 = h_c + h_r.$$

Assume a 200°F outside temperature. The Grashoff Number is given by:

$$\begin{aligned} G_r &= \frac{\rho^2 g\beta}{\mu^2}\left(T - T_a\right)\left(L^3\right) \\ &= \left(0.85 \times 10^6\right)\left(200 - 70\right)\left(40^3\right) \\ &= 7.1 \times 10^{12} \text{ : Turbulent free convection} \end{aligned}$$

$$P_r = 0.72$$

The average free convection Nusselt Number is given by:

$$\overline{N}_u = 0.13\left[G_r P_r\right]^{\frac{1}{3}}$$

$$= 0.13\left[\left(7.1\times10^{12}\right)\left(0.72\right)\right]^{\frac{1}{3}}$$

$$= 2240$$

$$= \frac{h_c L}{k}$$

$$k = 0.0161\frac{Btu}{hr - ft{-}^\circ F}$$

$\therefore$

$$h_c = \frac{(0.0161)(2240)}{40} \approx 1\frac{Btu}{hr - ft^2 - ^\circ F}$$

The radiant heat loss per ft of vessel is given by:

$$Q_r = seA\left(T^4 - T_a{}^4\right) = h_r A\left(T - T_a\right)$$

$\therefore$

$$h_r = \frac{se\left(T^4 - T_a{}^4\right)}{\left(T - T_a\right)}$$

$For\ e = 1$

$$h_r = \left(\frac{\left(0.171\times10^{-8}\right)\left(660^4 - 530^4\right)}{130}\right) = 1.45\ \frac{Btu}{hr - ft^2 - ^\circ F}$$

$\therefore$

$$h_o \approx 2.5\frac{Btu}{hr - ft^2 - ^\circ F}$$

Assume h_i is very large.

r_o= 12 in.
r_i = 5 in.
k = 0.1 Btu/hr-ft-°F

$\therefore$

$$\dot{Q} = \frac{(1560 - 70)}{\frac{\ln\left(\frac{12}{5}\right)}{(2\pi)(0.1)(40)} + \frac{1}{(2.5)(2\pi)(40)}} = 40700\frac{Btu}{hr}$$

This implies an actual outside wall temperature of:

$$T_o = T_i - \frac{\dot{Q}\ln\left(\frac{r_o}{r_i}\right)}{2\Pi kL} = 142°F$$

Since this is less than the originally assumed 200°F wall temperature, the calculated heat loss is conservative.

3.4 Tube Temperature Profile

The calciner tube temperatures will be monitored by attached thermocouples. The data collected will be used to better understand the self fluidized calcination process. Temperature data will also be used as an input variable of the control system. Some upset conditions could cause excessive heating and thermal expansion of the calciner tubes. If excessive temperatures are detected, corrective action will be taken before the system is damaged.

A precise determination of the tube temperature profile requires a numerical solution that considers the actual dehydration curve. Numerical solutions are done using ASPEN, a process simulation computer program. However, significant uncertainties remain including:

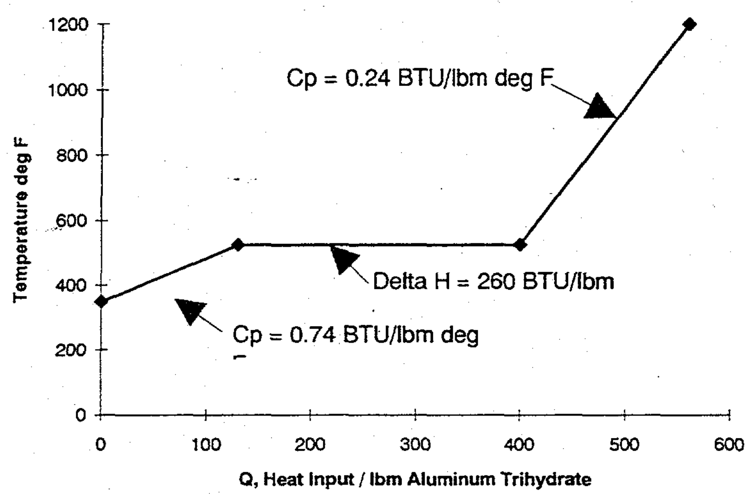
- Local variations of the inside and outside calciner tube film coefficients.
- Local porosity of the fluidized alumina hydrate.
- Local thermal equilibrium between the alumina hydrate and evolved steam.

Given these uncertainties, it is reasonable to simplify the dehydration curve to permit an analytical solution for the tube temperature profile. An analytical solution has two major advantages:

- It clearly shows the influence of various parameters.
- It provides a useful guide and check for the development of more advanced numerical solutions.

A simplified temperature vs. heat of dehydration curve is shown in Figure 11.

Figure 11: Simplified Dehydration Curve



The simplified dehydration curve implies that there are three distinct zones in the calciner tube as shown in Figure 12.

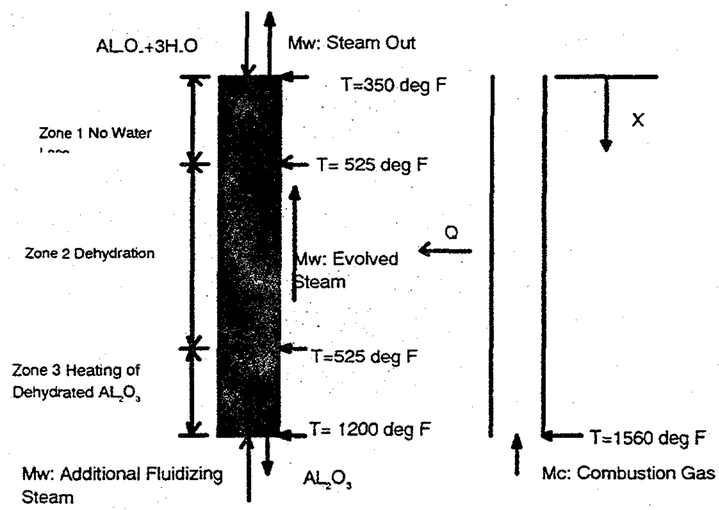


Figure 12: Simplified Calciner

Assuming the evolved steam and alumina are in local thermal equilibrium, we can formulate a two equation model; the unknowns are the alumina and flue gas temperatures. The governing equations for zones 1 and 3 are developed with the aid of the control volumes shown in Figure 13.

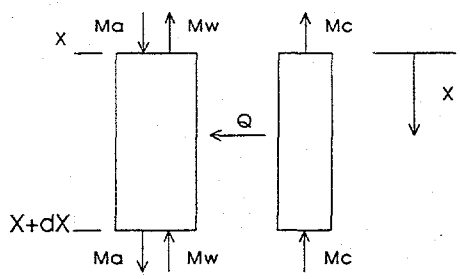


Figure 13. Calciner Control Volume

Where:

$\dot{m}_a$ = alumina flow rate
$\dot{m}_w$ = steam flow rate
$\dot{m}_c$ = combustion gas flow rate

Because there is no evolution of steam in zones 1 and 3, from conservation of mass we have:

$$\frac{d\dot{m}_a}{dx} = 0$$
$$\frac{d\dot{m}_w}{dx} = 0$$
$$\frac{d\dot{m}_c}{dx} = 0$$

For steady-state conditions, the first law is:

$$\dot{Q} = \sum \dot{m}_e h_e - \sum \dot{m}_i h_i$$

For the fluidization tube we have:

$$\dot{Q} = \dot{m}_a \left(h_{x+Dx} - h_x\right)_a + \dot{m}_w \left(h_x - h_{x+Dx}\right)_w$$

let $Dh = C_p DT_a$

$\therefore$

$$\dot{Q} = \dot{m}_a C_{pa} \left(T_{x+Dx} - T_x\right)_a + \dot{m}_w C_{pw} \left(T_x - T_{x+Dx}\right)_a$$

$$\therefore$$

$$\dot{Q} = \left(\dot{m}_a C_{pa} - \dot{m}_w C_{pw}\right) DT_a$$

For the flue gas :

$$-\dot{Q} = \dot{m}_c C_{pc}\left(T_x - T_{x+Dx}\right)_c$$
$$-\dot{Q} = -\dot{m}_c C_{pc} DT_c$$
$$\therefore$$
$$\dot{Q} = \dot{m}_c C_{pc} DT_c$$

We also have the phenomenological equation:

$$\dot{Q} = UdA\left(T_c - T_a\right)$$
$$U \approx h_o$$
$$dA = \pi D dx$$
$$\therefore$$
$$\dot{Q} = \pi Dh\left(T_c - T_a\right)dx$$

In zones 1 and 3, the alumina and combustion gas temperatures are governed by:

$$\frac{dT_a}{dx} = \frac{pDh_o}{\left[\dot{m}_a C_{pa} - \dot{m}_w C_{pw}\right]}\left(T_c - T_a\right)$$
$$\frac{dT_c}{dx} = \frac{pDh_o}{\dot{m}_c C_{pc}}\left(T_c - T_a\right)$$

Let:

$$a_1 = \frac{pDh_o}{\left[\dot{m}_a C_{pa} - \dot{m}_w C_{pw}\right]}$$
$$a_2 = \frac{pDh_o}{\dot{m}_c C_{pc}}$$
$$\therefore$$
$$\frac{dT_a}{dx} = -a_1 T_a + a_1 T_c$$
$$\frac{dT_c}{dx} = -a_2 T_a + a_2 T_c$$

To find the general solution let:

$$T_a = Ae^{\lambda x}$$
$$T_c = Be^{\lambda x}$$
$$\therefore$$
$$A\lambda e^{\lambda x} = -a_1 Ae^{\lambda x} + a_1 Be^{\lambda x}$$
$$B\lambda e^{\lambda x} = -a_2 Ae^{\lambda x} + a_2 Be^{\lambda x}$$
$$\therefore$$
$$(a_1 + \lambda)A - a_1 B = 0$$
$$a_2 A + (-a_2 + \lambda)B = 0$$
$$\therefore$$
$$\lambda\left[\lambda + (a_1 - a_2)\right] = 0$$

This last equation has two solutions:

$$\lambda_1 = 0$$
$$\lambda_2 = -(a_1 - a_2)$$

$\therefore$

For $\lambda_1 = 0$

$A_1 = B_1 = 1$

For $\lambda_2 = -(a_1 - a_2)$

$\therefore$

$a_2 A - a_1 B = 0$

Let $A_2 = 1$

$\therefore$

$$B_2 = \frac{a_2}{a_1}$$

$\therefore$

$$T_a = C_1 + C_2 e^{-(a_1 + a_2)x} \qquad \text{Ref.3}$$
$$T_c = C_1 + C_2 \frac{a_2}{a_1} e^{-(a_1 - a_2)x}$$

at $x = 0$ $T_a = T_{ai}$

at $x = L$ $T_c = T_{ci}$

The general solution for the combustion gas temperature in zones 1 and 3 is:

$$T_a = T_{ai} + \left\{\frac{T_{ci} - T_{ai}}{1 - \frac{a_2}{a_1}e^{-(a_1-a_2)L}}\right\}\left\{1 - e^{-(a_1-a_2)x}\right\}$$

$$T_c = T_{ai} + \left\{\frac{T_{ci} - T_{ai}}{1 - \frac{a_2}{a_1}e^{-(a_1-a_2)L}}\right\}\left\{1 - \frac{a_2}{a_1}e^{-(a_1-a_2)x}\right\}$$

In zone 2 the alumina temperature is constant, $T_a = 525°F$. The heat transfer required from the flue gas to the hydrated alumina is $Dh = 260 \frac{Btu}{lbm}$ of alumina hydrate. The total heat transfer required is $\dot{m}_a Dh$. The total change in temperature of the flue gas is:

$$\Delta T_c = \frac{\dot{m}_a \Delta h}{\dot{m}_c C_{pc}}$$

The local change in flue gas temperature is governed by:

$$\frac{dT_c}{dx} = \frac{\pi Dh}{\dot{m}_c C_{pc}}\left(T_c - T_a\right)$$

$\therefore$

$$T_c = T_a + Ce^{\frac{\pi Dh}{\dot{m}_c C_{pc}}x}$$

To find the constant C, we will arbitrarily impose the following boundary condition. The rational will be explained in the discussion of the general solution procedure.

$At\, x = 0$

$T_c = T_{c2}$

$\therefore$

$C = T_{c2} - T_a$

$\therefore$

$$T_c = T_a + \left(T_{c2} - T_a\right)e^{\frac{pDh}{\dot{m}_c C_{pc}}x}$$

The general the two equation thermal model of the calciner is summarized by equations 1, 2, and 3 shown below.

$$1)\ T_a = T_{ai} + \left\{ \frac{T_{ci} - T_{ai}}{1 - \frac{a_2}{a_1} e^{-(a_1 - a_2)L}} \right\} \left\{ 1 - e^{-(a_1 - a_2)x} \right\}$$

$$2)\ T_c = T_{ai} + \left\{ \frac{T_{ci} - T_{ai}}{1 - \frac{a_2}{a_1} e^{-(a_1 - a_2)L}} \right\} \left\{ 1 - \frac{a_2}{a_1} e^{-(a_1 - a_2)x} \right\}$$

$$3)\ T_c = T_a + \left(T_{c2} - T_a\right) e^{\frac{pDh}{\dot{m}_c C_{pc}} x}$$

Consider Figure 14. There are six unknowns shown:

- The three lengths l_1, l_2, and l_3.
- The intermediate combustion gas temperatures, T_{c1}, and T_{c2}.
- The flue gas outlet temperature, T_{co}.

We can use the following solution procedure:

1. Use an overall energy balance to determine the intermediate and outlet flue gas temperatures.

2. With the combustion gas temperatures known, the Log Mean Temperature Difference and lengths for the three zones can be computed.

3. Use equations 1 through 3 to compute the alumina and combustion gas temperatures as a function of length.

4. The two equation thermal model is applied to a three zone calciner model in which the alumina specific heat changes between zones. We will make two additional assumptions:

5. There is no additional fluidizing steam.

6. The mass flow rate of calcined alumina is equal to the inlet flow rate of alumina trihydrate. This assumption requires a bit of explanation. The alumina specific heat values in zones 1 and 3 are determined from the heat input vs. temperature curves shown in Figures 1 and 10 that are normalized on one pound mass of initial alumina trihydrate. Changes in mass due to evolution of steam are not considered. This is of no importance because the mass flow rate

and specific heat never occur as separate terms. They always appear in the combination, $\dot{m}_a C_{pa}$, and this product will always have the correct value under this assumption.

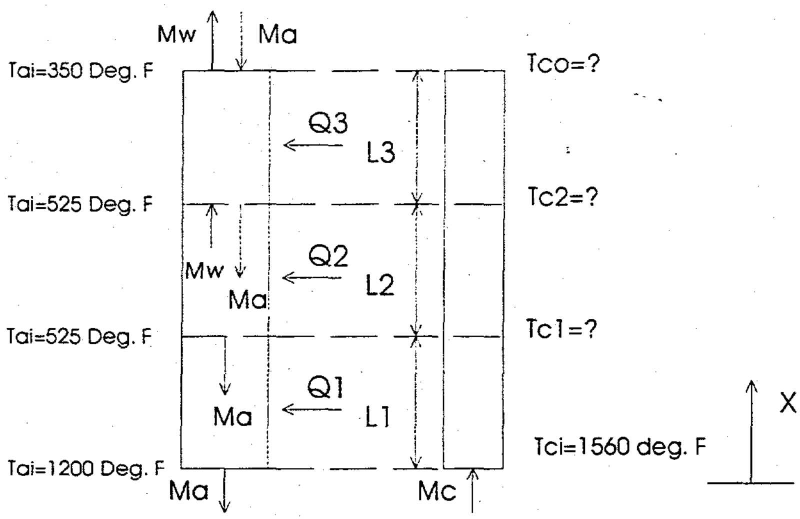


Figure 14: Thermal Model for Simplified Dehydration Curve

Figures 15 through 18 compare results of the simplified model with ASPEN numerical solutions.

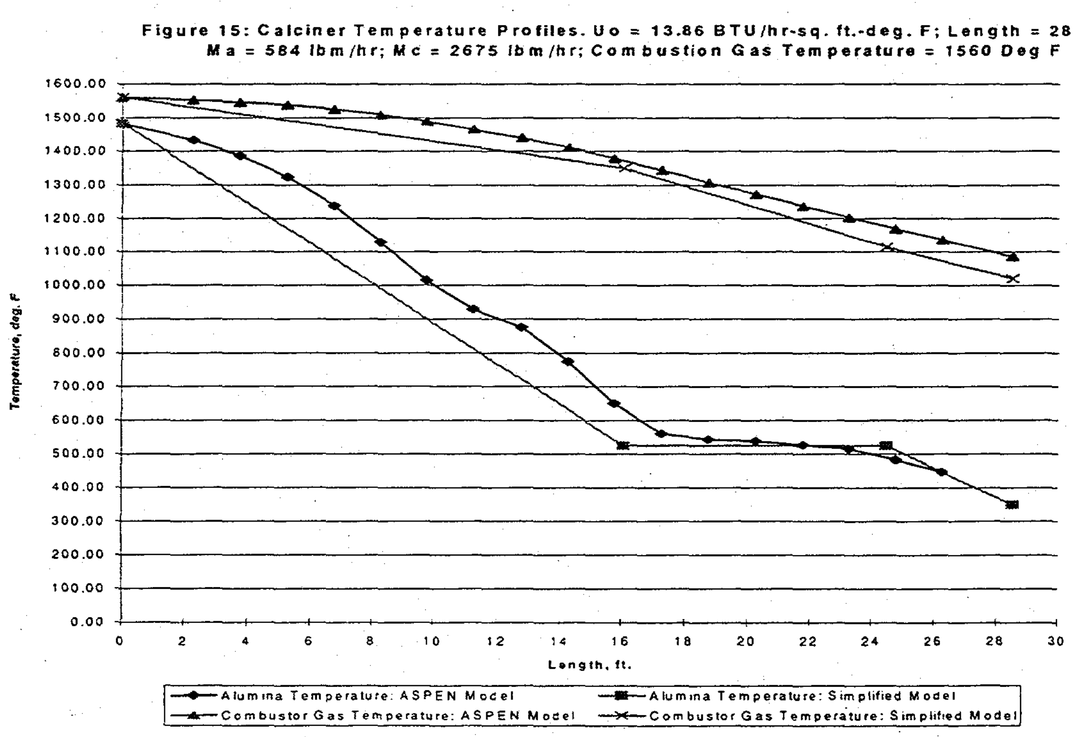


Figure 16: Calciner Temperature Profiles. Uo = 13.86 BTU/hr-sq. ft.-deg. F; Length = 27 ft.; Ma = 585 lbm/hr;Mc = 2675 lbm/hr; Combustion Gas Temperature = 1560 Deg F

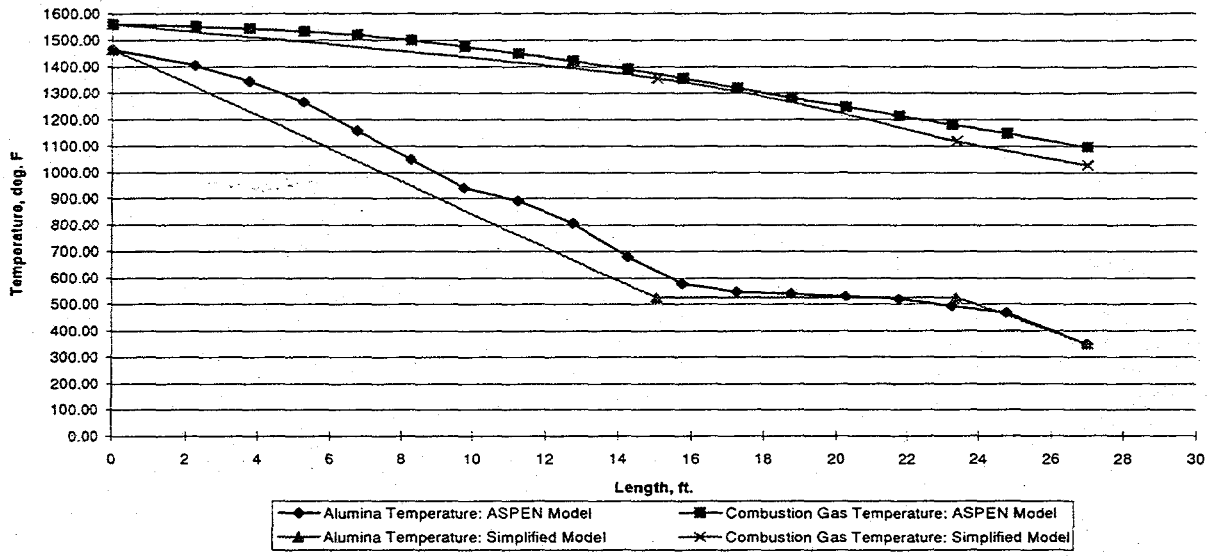


Figure 17: Calciner Temperature Profile. Uo = 13.86 BTU/hr-sq. ft.-deg. F; Length = 27 ft.; Ma = 584 lbm/hr; Mc = 2675 lbm/hr; Combustion Gas Temperature = 1472 Deg F

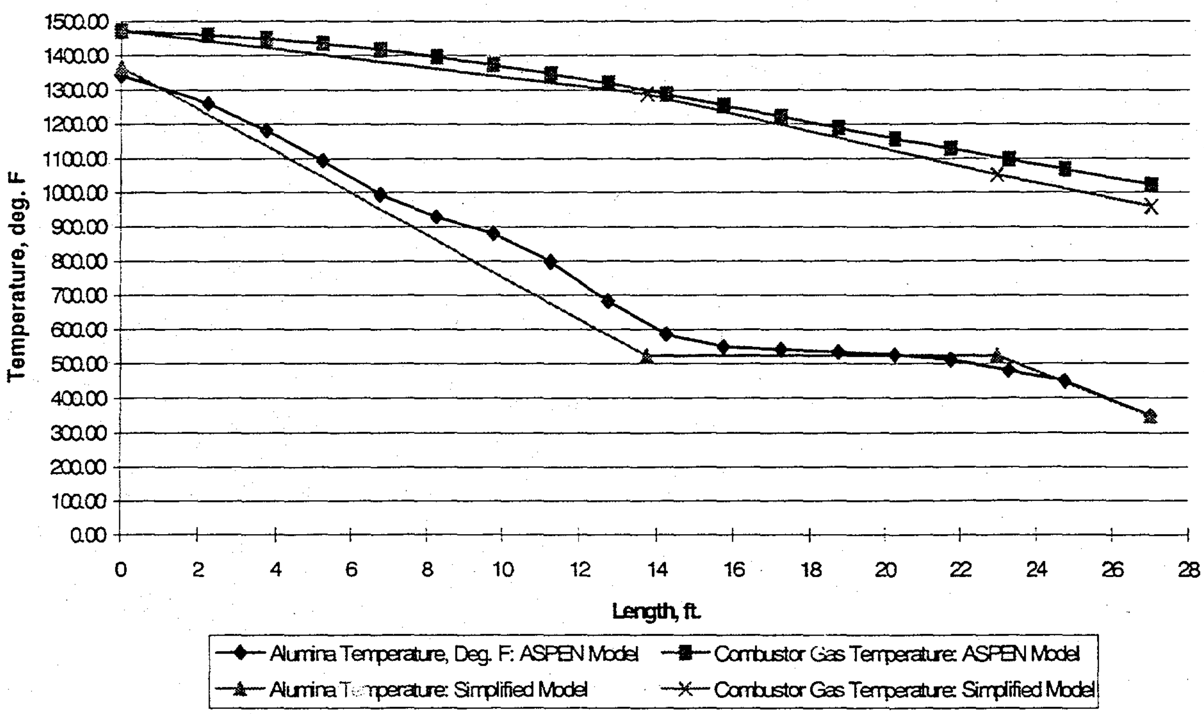
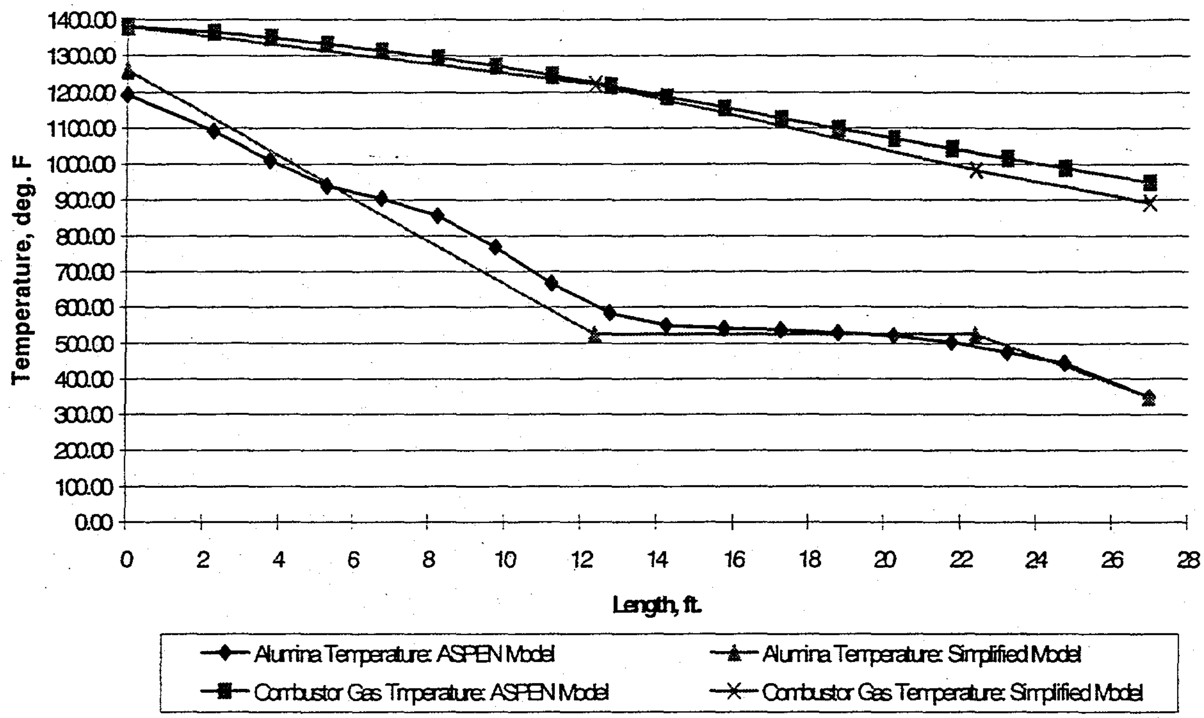


Figure 18: Calciner Temperature Profile. Uo = 13.86 BTU/hr-sq. ft.-deg. F; Length = 27 ft.; Ma = 584 lbm/hr; Mc = 2675 lbm/hr; Combustion Gas Temperature = 1382 Deg F



There is very good agreement between the results of the simplified model and the ASPEN numerical solutions, and we will use it to perform preliminary parameter studies.

4.0 Tube Thermal Expansion

Differential expansion between the calciner tubes and shell is accommodated by individual expansion joints installed at one end of each tube. The carbon steel shell is internally insulated and will operate at approximately 200°F. In a worst case upset condition, a plugged calciner tube could approach a uniform temperature equal to the maximum combustion gas inlet temperature, 1560°F. The mean coefficients of thermal expansion for carbon and stainless steels are approximately:

$$\alpha_{carbon\ \mathrm{steel}} \approx 6.5 \times 10^{-6} \frac{in}{in-{}^\circ F}$$

$$\alpha_{stainless\ \mathrm{steel}} \approx 10.5 \times 10^{-6} \frac{in}{in-{}^\circ F}$$

The worst case thermal differential expansion is approximately:

$$DL = (27\ ft.)\left(12\ \frac{in}{ft}\right)\left[(1560)\left(10.5\times 10^{-6}\right)-(200)\left(6.5\times 10^{-6}\right)\right]=4.89\ in$$

The expansion joints are designed to be installed with a precompression of 3 in. This is approximately the maximum thermal expansion we would expect under normal operating conditions. If the tube expands 3 in., the only operating condition stresses are due to differential pressure. A greater expansion will put the joints in axial tension. This condition will not result in immediate failure, but will reduce the cyclic lifetime of the joints. We wish to avoid this condition and will use the simplified thermal model to examine a few potential cases. We expect the simplified thermal model will be quite useful as we refine the system control parameters.

We can control the combustion gas flow rate through the calciner and the combustion gas inlet temperature. Flow rates control heat transfer coefficients to some extent. However, the heat transfer coefficients still present a significant uncertainty in the analysis. We can approximate the tube external heat transfer coefficient using reasonably well established correlations and experience with similar applications. We have little experience with conditions on the fluidized side of the tube. The initial calculations were based on the assumption that this heat transfer coefficient will be very large. This is the probable case but we should consider a more conservative condition that results in greater tube expansion relative to the calciner shell. The overall heat transfer coefficient is given by:

$$U_o = \frac{1}{\dfrac{r_o}{r_i}\dfrac{1}{h_i}+\dfrac{r_o \ln\left(\dfrac{r_o}{r_i}\right)}{k}+\dfrac{1}{h_o}}$$

Consider the tube temperature profiles for the conditions shown in Table 4.

Table 4: Tube Overall Conductance

r_i in	r_o in	k BTU/hr-ft-°F	h_i BTU/hr-ft^2-°F	h_o BTU/hr-ft^2-°F	U_o BTU/hr-ft^2-°F
1.034	1.188	11	∞	14	13.76
1.034	1.188	11	28	14	8.79

The tube temperature profiles for these conditions are shown in Figures 19 and 20.

Figure 19: Calciner Temperature Profile. Uo = 13.76 BTU/hr-sq.-ft.-deg. F; Length = 27 ft.; Ma = 584 lbm/hr; Mc = 2765 lbm/hr; Combustion Gas Temperature = 1337 deg. F

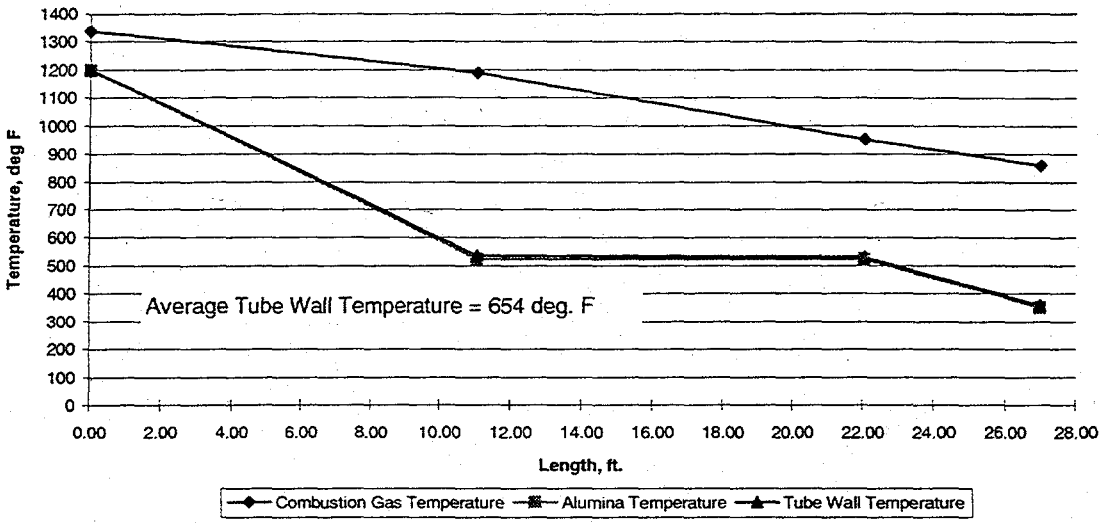
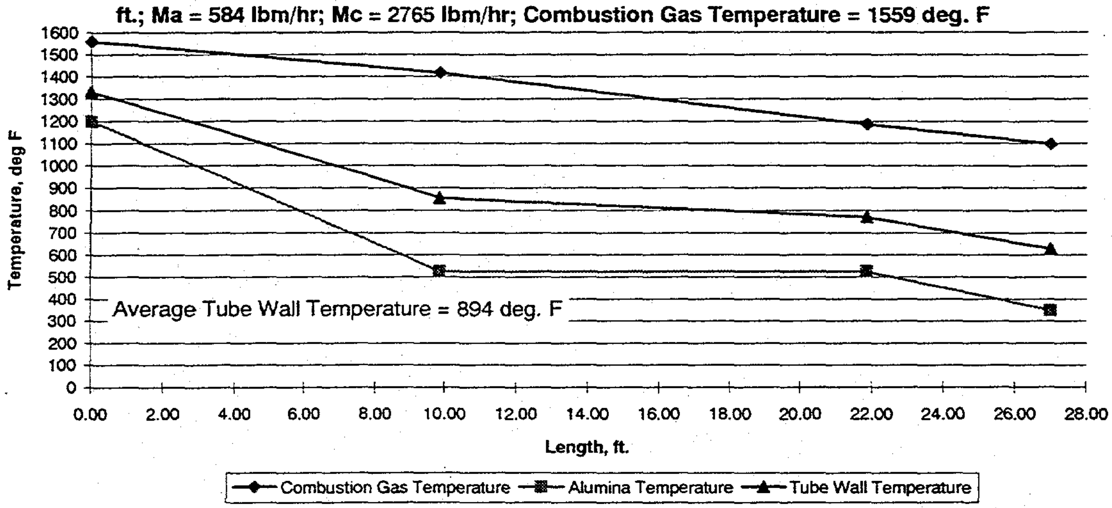


Figure 20: Calciner Temperature Profile. Uo = 8.80 BTU/hr-sq.-ft.-deg. F; Length = 27 ft.; Ma = 584 lbm/hr; Mc = 2765 lbm/hr; Combustion Gas Temperature = 1559 deg. F



The tube thermal expansions for the two cases are:

$$\Delta L = (27\,ft.)\left(12\ \frac{in}{ft}\right)\left[(654)\left(10.5\times10^{-6}\right)-(200)\left(6.5\times10^{-6}\right)\right] = 1.70\ in$$

$$\Delta L = (27\,ft.)\left(12\ \frac{in}{ft}\right)\left[(894)\left(10.5\times10^{-6}\right)-(200)\left(6.5\times10^{-6}\right)\right] = 2.62\ in$$

5.0 Recycle Ratio

The recycle ratio depends on the combustor flame temperature, combustor excess air and total heat loss from the combustion gas flowing through the calciner. The total heat loss includes heat supplied to the alumina, losses through the vessel and piping insulation, and possible heat supplied to the alumina preheater. This latter quantity becomes important if the recouperator supplying hot air to the alumina preheater is placed before the combustion gas recycle diverter valve. Since this effect is expected to be small, we will neglect it for the moment. The total heat loss considered is:

$$\dot{Q} = 340{,}000 + 40700 = 380{,}700 \ \frac{Btu}{hr}$$

Consider a control volume enclosing the combustion gas on the shell side of the calciner, the recycle compressor, and the mixer as shown in Figure 21.

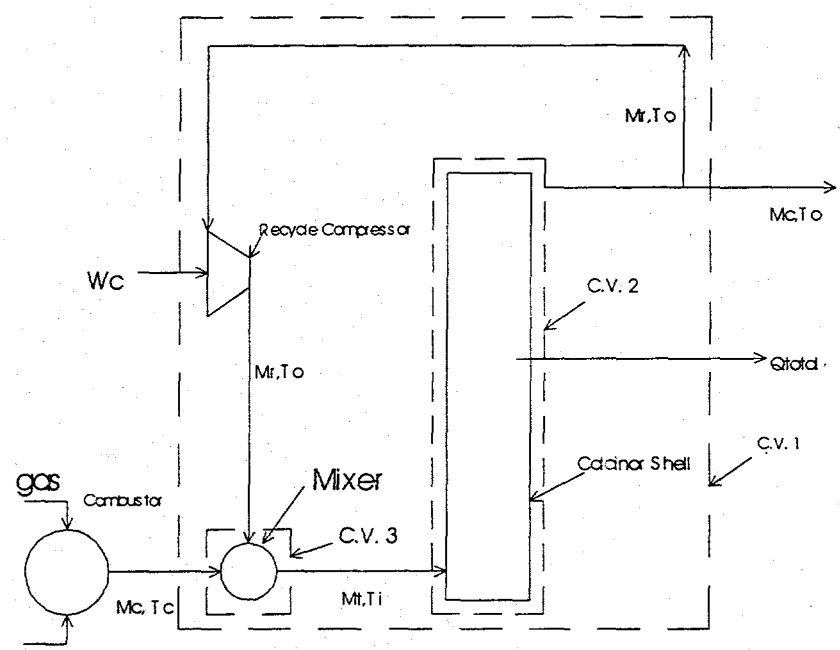


Figure 21: Calciner Thermal Model

Applying the first law to control volume 1 we have:

$$\dot{Q} = \dot{m}_c C_p \left[T_c - T_o \right]$$

For a given product throughput, $\dot{Q}$ is known.

$$\dot{Q} = 380{,}700 \ \frac{Btu}{hr} \ \text{for } 265 \ \frac{kg}{hr} \ \text{of } Al_2O_3 \cdot 3H_2O$$

For a given fuel/air ratio, T_c is known. T_c for three fuel air ratios is calculated as follows.

Combustor Temperature for Theoretical Air

$$CH_4 + 2O_2 + (2)(3.76)N_2 \rightarrow CO_2 + 2H_2O + 7.52N_2$$

First Law:

$$H_R = H_P$$

$$\sum_R n_i\left[\bar{h}_f^o + Dh\right]_i = \sum_P n_e\left[\bar{h}_f + D\bar{h}\right]_e$$

Assuming ideal gas behvior, at 77° F we have:

$$H_R = -32{,}211 \frac{Btu}{lb\text{-}mole}$$

$$H_P = -377369 + D\bar{h}_{CO_2} + 2D\bar{h}_{H_2O} + 7.52D\bar{h}_{N_2}$$

Temp °F	$\sum H_P^0$	$\Delta\bar{H}_{CO_2}$	$2\Delta\bar{H}_{H_2O}$	$7.52\Delta\bar{H}_{N_2}$	$\sum H_P$
4140	-377369	47182	75970	216779	-37438
4320	-377369	49813	80540	228570	-18446

By interpolation:

T = 4190 °R ≈ 3700 °F

Combustor Temperature for 150% Theoretical Air

$$CH_4 + (2)(1.5)O_2 + (2)(3.76)(1.5)N_2 \rightarrow CO_2 + 2H_2O + O_2 + 11.28N_2$$

Temp °F	$\sum H_P^0$	$\Delta\bar{H}_{CO_2}$	$2\Delta\bar{H}_{H_2O}$	$\Delta\bar{H}O_2$	$11.28\Delta\bar{H}_{N_2}$	$\sum H_P$
3240	-377369	34117	53870	22237	237703	-29382
3060	-377369	31617	49634	20637	220456	-55025

By interpolation:

T = 3220 °R = 2760 °F

Combustor Temperature for 300% Theoretical Air

$$CH_4 + (2)(3)O_2 + (2)(3.76)(3)N_2 \rightarrow CO_2 + 2H_2O + 4O_2 + 22.56N_2$$

Temp °F	$\sum H_P^0$	$\Delta\bar{H}_{CO_2}$	$2\Delta\bar{H}_{H_2O}$	$4\Delta\bar{H}O_2$	$22.56\Delta\bar{H}_{N_2}$	$\sum H_P$
2160	-377369	19138	29664	51220	272796	-4551
1980	-377369	16733	25956	45116	240287	-49277

By interpolation:

T = 2049 °R = 1590 °F

For an assumed outlet temperature, $\dot{m}_c$ can be calculated.

Applying the first law to control volume 2 we have:

$$\dot{Q} = \dot{m}_t C_p [T_i - T_o]$$

$\therefore$

$$\dot{m}_t = \frac{\dot{Q}}{C_p (T_i - T_o)}$$

The ratio of total mass flow rate to combustor gas flow rate is:

$$\frac{\dot{m}_t}{\dot{m}_c} = \frac{[T_c - T_o]}{[T_i - T_o]}$$

where $\dot{m}_t = \dot{m}_c + \dot{m}_r$

The ratio of recycle flow rate to combustor flow rate is:

$$\frac{\dot{m}_r}{\dot{m}_c} + 1 = \frac{[T_c - T_o]}{[T_i - T_o]}$$

$\therefore$

$$\frac{\dot{m}_r}{\dot{m}_c} = \frac{[T_c - T_i]}{[T_i - T_0]}$$

We can check these equations by applying the first law to control volume 3 around the mixer. For steady state adiabatic mixing:

$$\sum_e \dot{m}_e h_e = \sum_i \dot{m}_i h_i$$

$$\therefore$$

$$\dot{m}_c C_p T_c + \dot{m}_r C_p T_o = \dot{m}_t C_p T_i$$

$$\dot{m}_t = \dot{m}_c + \dot{m}_r$$

$$\therefore$$

$$\dot{m}_c\left[T_c - T_i\right] = \dot{m}_r\left[T_i - T_o\right]$$

$$\therefore$$

$$\frac{\dot{m}_r}{\dot{m}_c} = \frac{\left[T_c - T_i\right]}{\left[T_i - T_o\right]}$$

$$\therefore$$

$$\frac{\dot{m}_t - \dot{m}_c}{\dot{m}_c} = \frac{\left[T_c - T_i\right]}{\left[T_i - T_o\right]}$$

$$\therefore$$

$$\frac{\dot{m}_t}{\dot{m}_c} = \frac{\left[T_c - T_o\right]}{\left[T_i - T_o\right]}$$

Plots of total mass flow rate, combustor mass flow rate, and the flow ratios $\frac{\dot{m}_t}{\dot{m}_c}$ and $\frac{\dot{m}_r}{\dot{m}_c}$ are shown in Figures 22 through 25.

Figure 22 : Combustor Flow Rate vs. Calcinor Temperature Drop

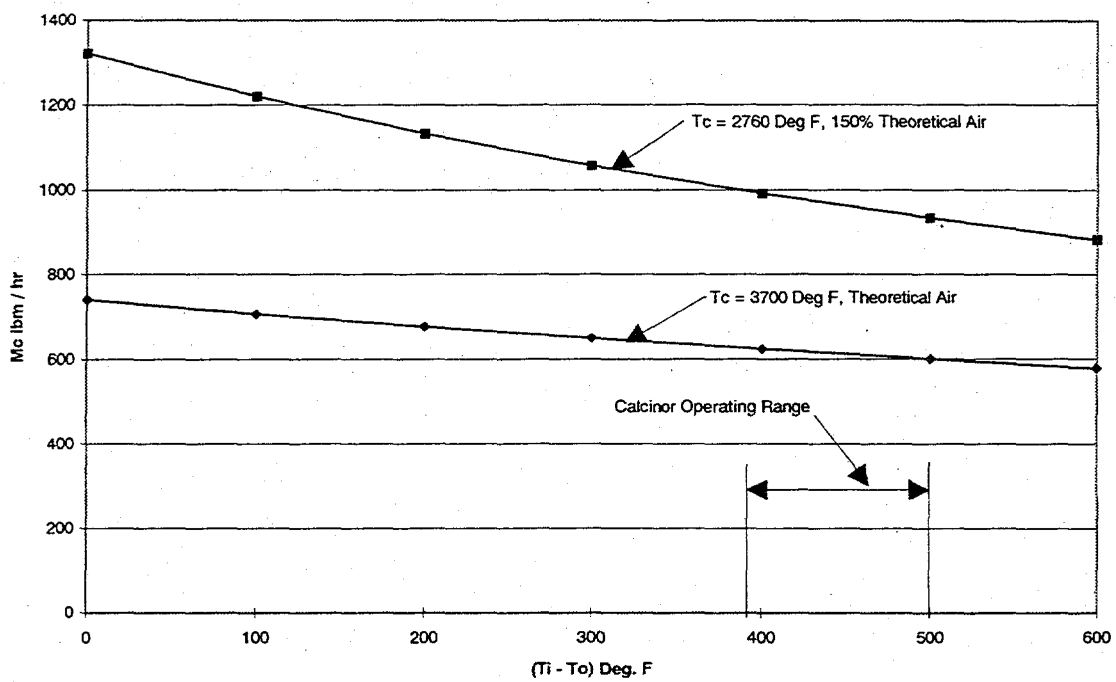


Figure 23 : Total Shell Side Gas Flow Rate vs. Calcinor Temperature Drop

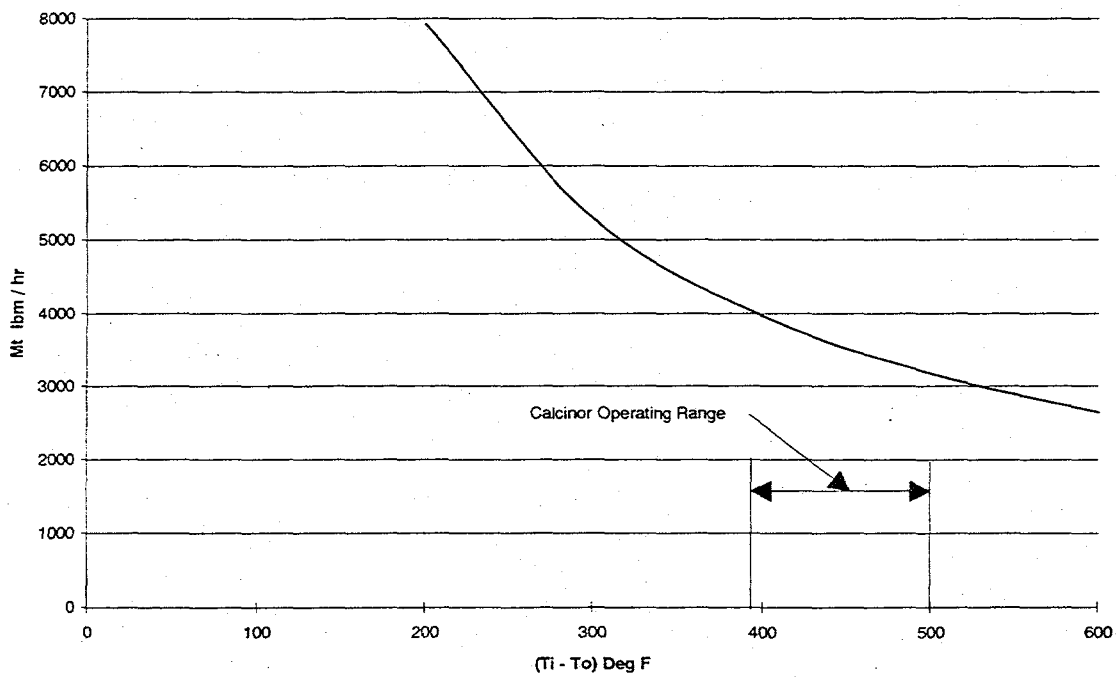


Figure 24 : Ratio of Total Shell Side Gas Flow Rate to Combustor Gas Flow Rate vs. Calcinor Temperature Drop

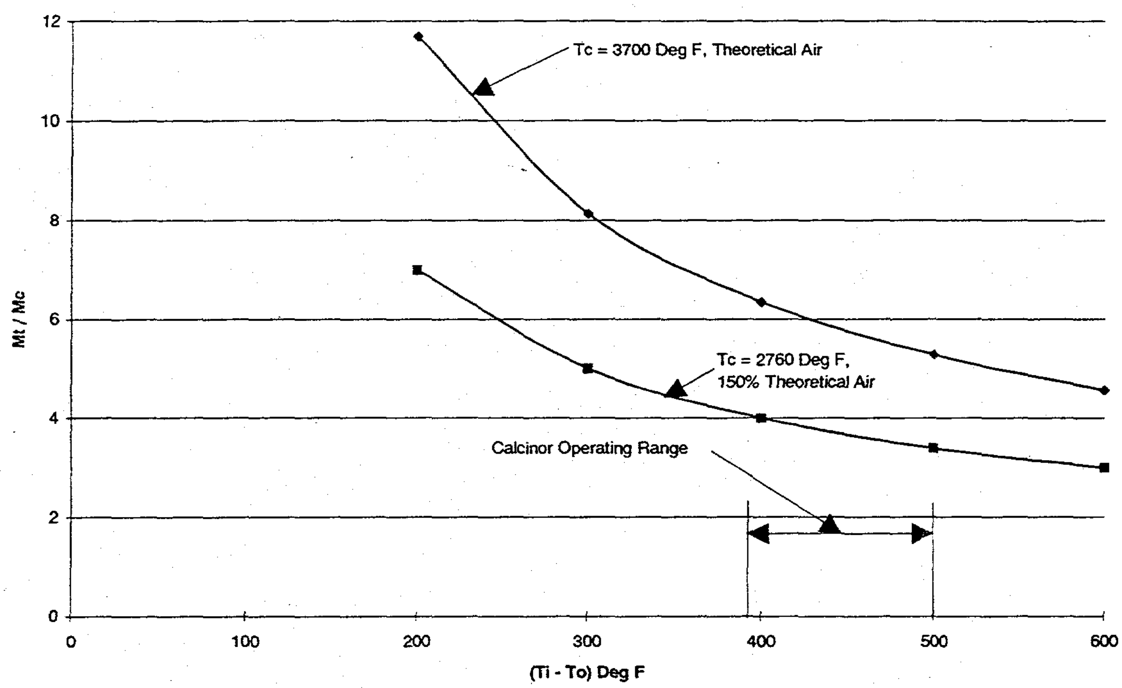
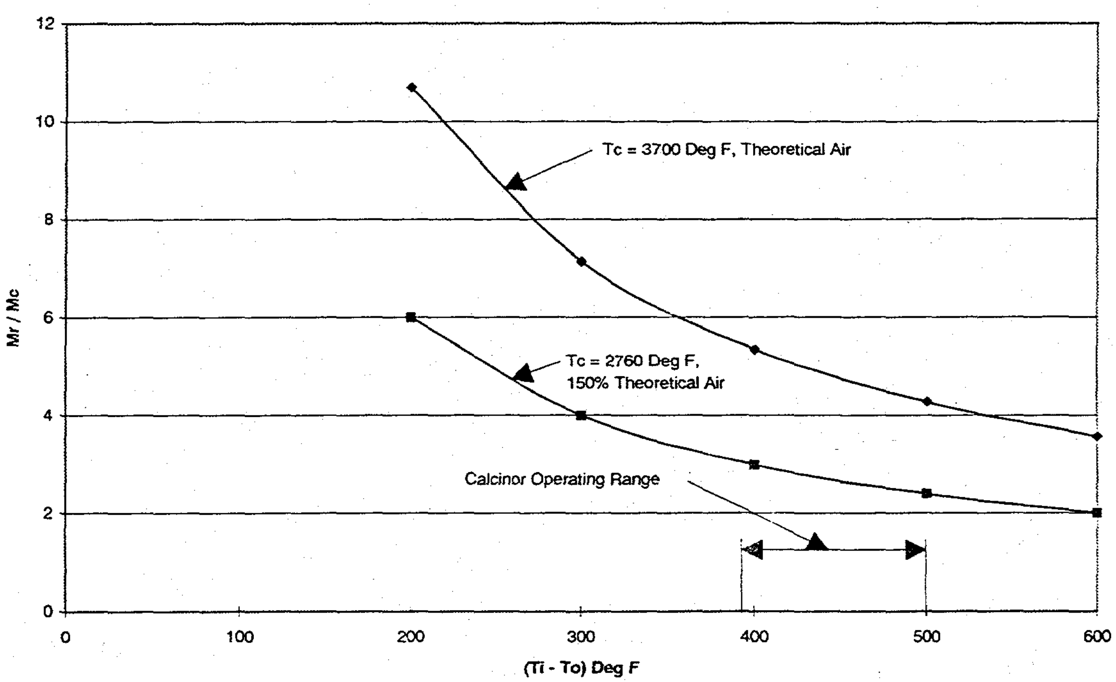


Figure 25 : Ratio of Recyle Flow Rate to Combustor Gas Flow Rate vs. Calcinor Temperature Drop



The pressurized combustion system is designed to operate with a maximum of 150% theoretical air (50% excess air). However, we expect that normal operation will use 10% excess air. With 150% excess air and a 390°F combustion gas temperature drop across the calciner, the combustor and total flow rates are:

$$\dot{m}_c = 998\frac{lbm}{hr}$$

$$\dot{m}_t = 4067\frac{lbm}{hr}$$

$$\therefore$$

$$\dot{m}_r = 4067 - 998 = 3069\frac{lbm}{hr}$$

$$\therefore$$

$$\frac{\dot{m}_r}{\dot{m}_c} = \frac{3069}{998} \approx 3$$

The worst case stack loss is:

$$\begin{aligned}\dot{Q}_{stack} &= \dot{m}_c C_{pc}\left(T_c - T_o\right) \\ &= (998)(.24)(2760 - 1170) \\ &= 380837\frac{BTU}{hr}\end{aligned}$$

The worst case combustor heat rate, $\dot{Q}_c$, is:

$$\dot{Q}_c = 340000 + 40700 + 380837 = 761537\frac{Btu}{hr}$$

The process efficiency is:

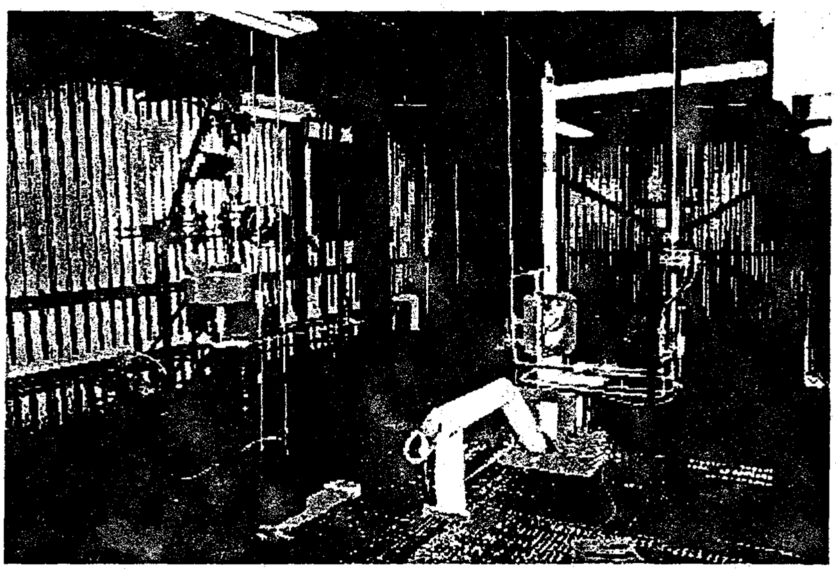
$$\eta = \frac{340000}{761537} = 45\%$$

6.0 References

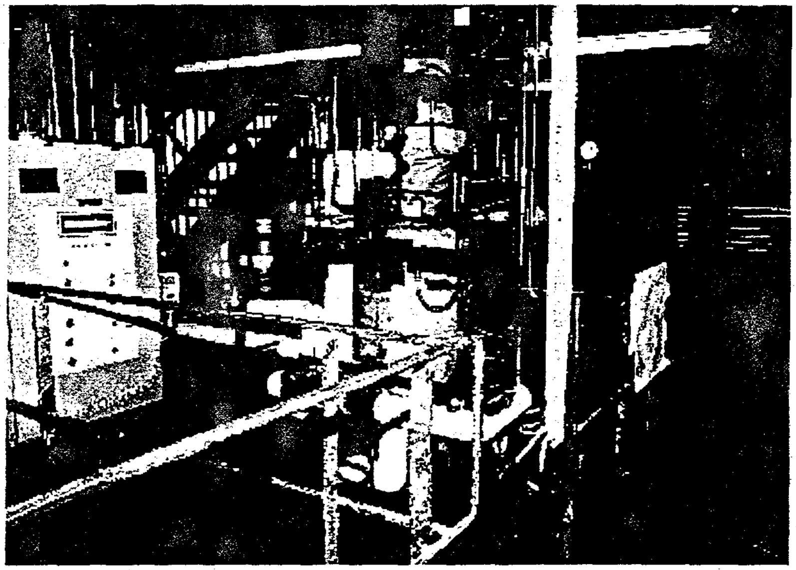
1. Heat Transfer 4th. ed., Holman, J. P., McGraw-Hill Book Company, 1976, New York.
2. Applied Fluid Dynamics Handbook, Blevins, R. D., 1984, Van Nostrand Reinhold Company, New York.
3. Advanced Calculus, Kaplan, W., 1959, Addison-Wesley Publishing Co., Reading, MA.

APPENDIX D

Picture No. 1: Hydrate Supper Sacks (Doubles as feed tank)
1.45 ton of material per sack
Location — Level 5

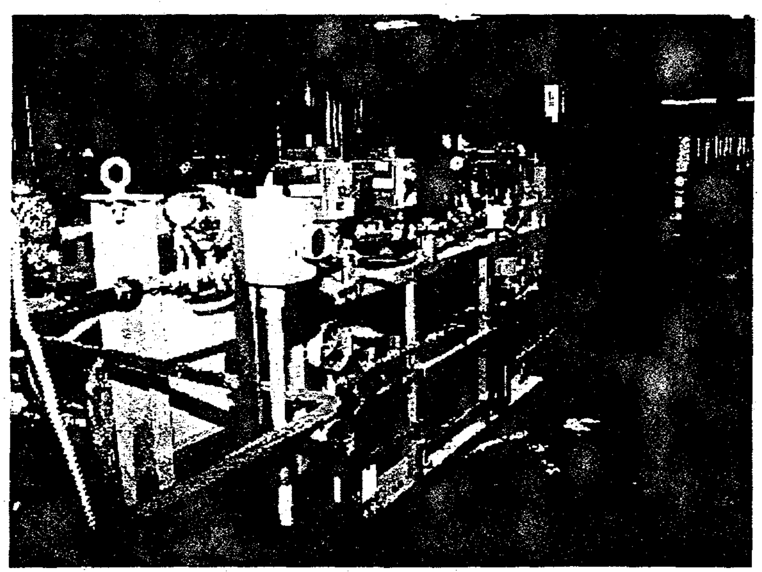


Picture No. 2: Combustor Pressure Control and Vent (Left Side)
Solids Feed Line (Center) and Steam Line to Condenser (Right Center)
Location — Level 5



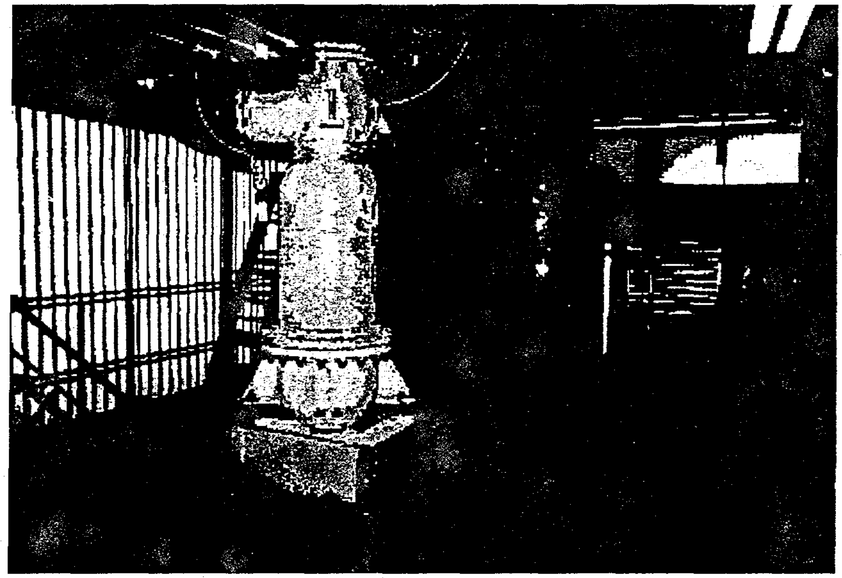
Picture No. 3: Macawber Feed System

Location — Level 4



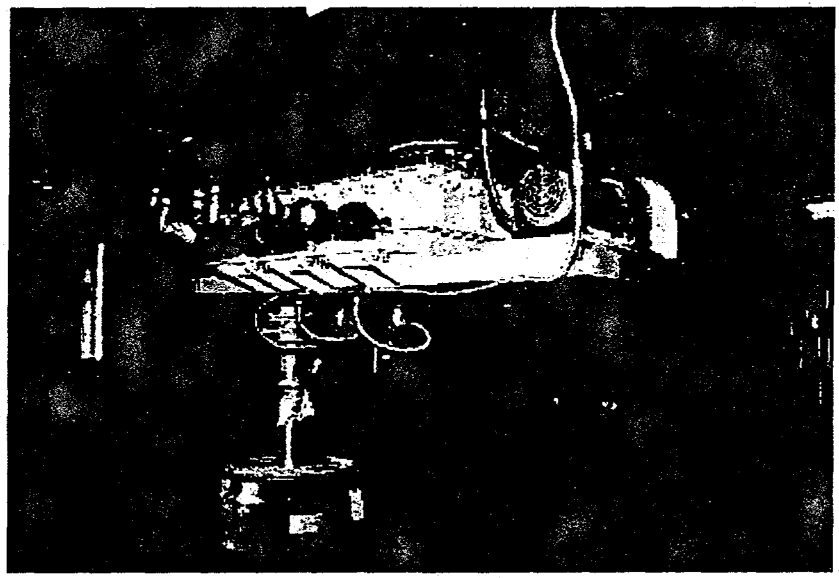
Picture No. 4: Combustor Gas Train
Air (Left Unit) Gas (Center)

Location — Level 3



Picture No. 5: Pulse Combustor (Gray Unit)
Calciner Vessel (Brown Unit)
Location — Level 2

Picture No. 6: Fluidization Injection System (Left Center)
Holding Section (Insulated Vessel) GEMCO Valves (Below Holding Vessel)
Location — Level 1

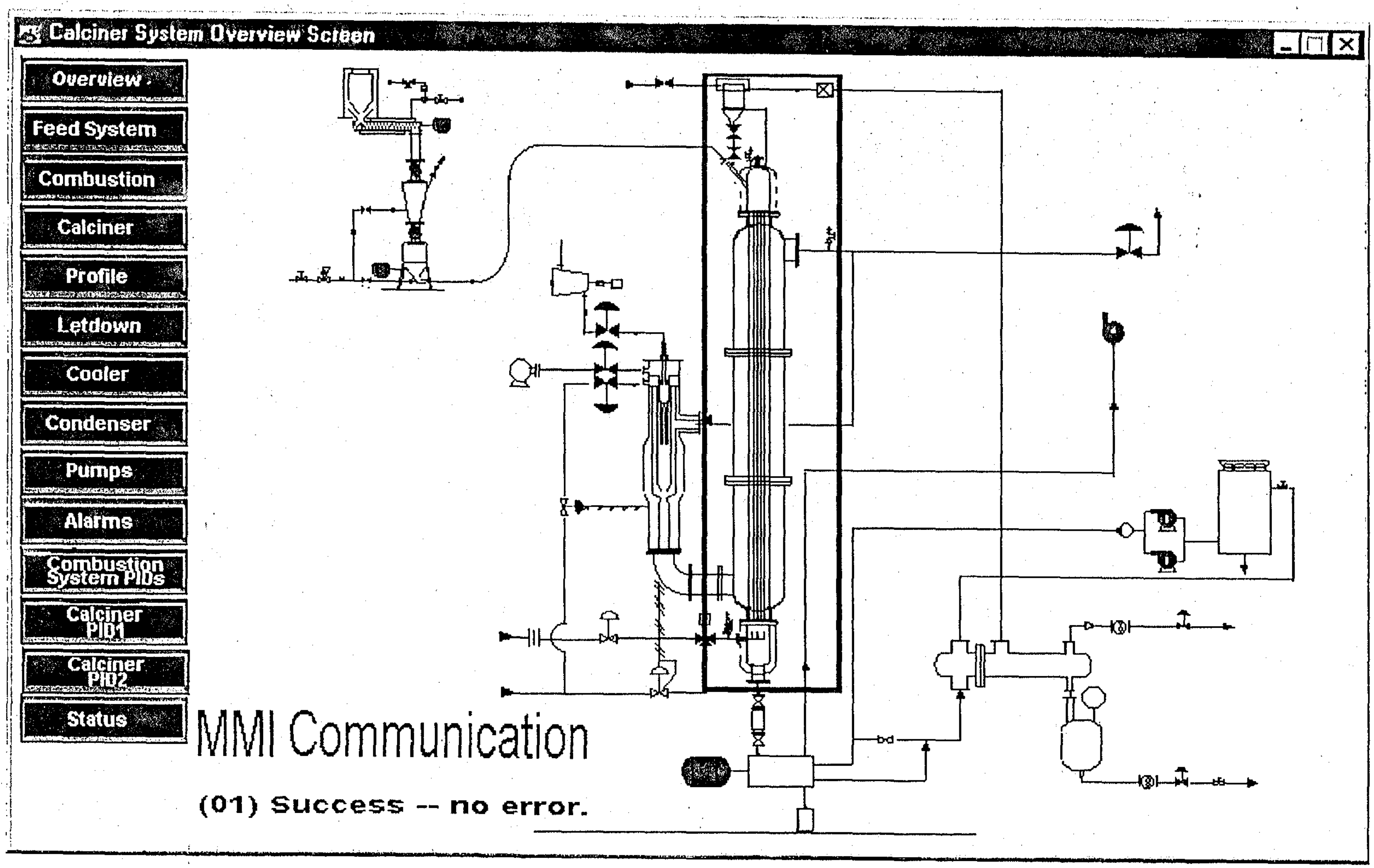


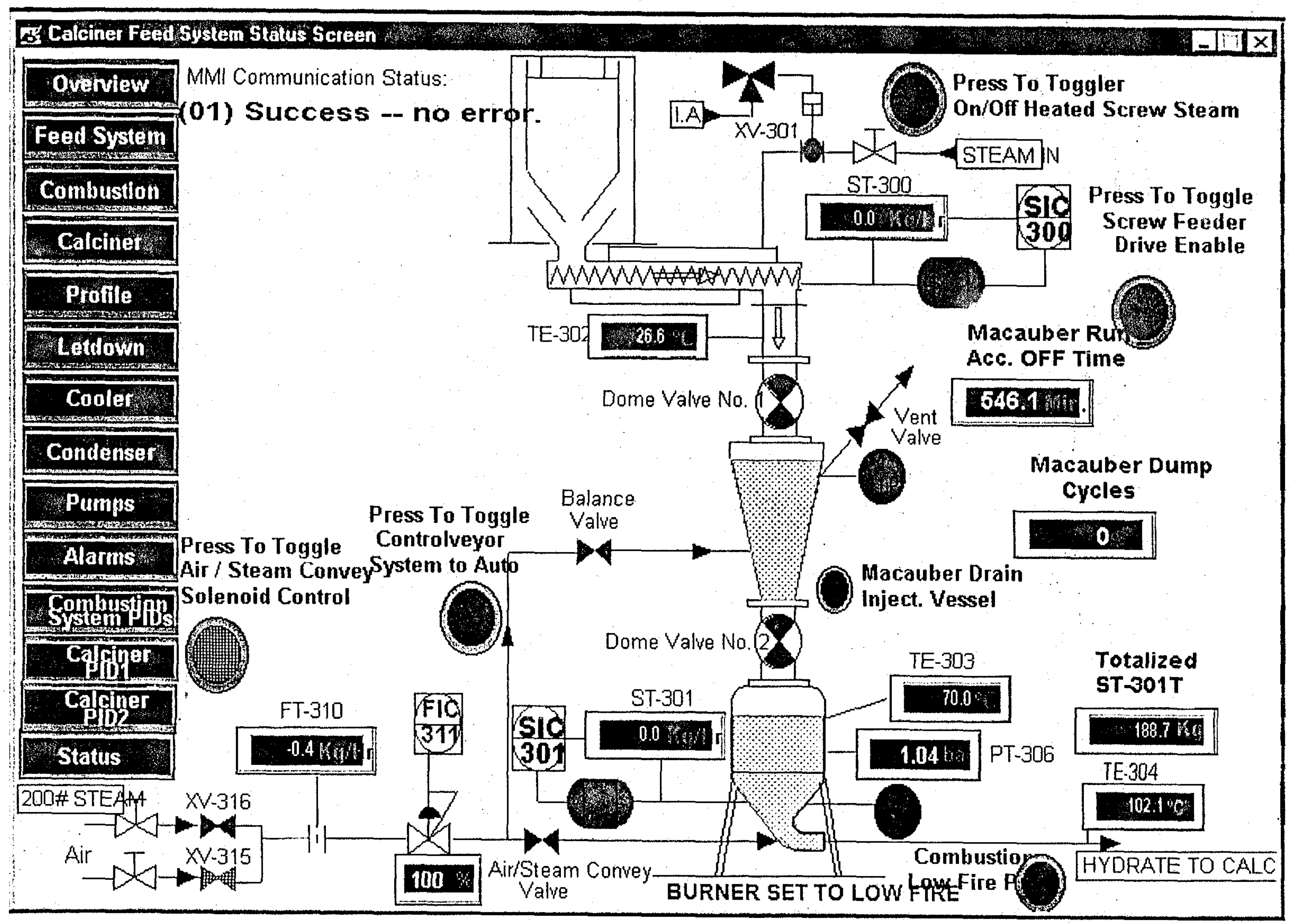
Picture No. 7: Disk Cooler and Product Tank (55 Gal Drum)

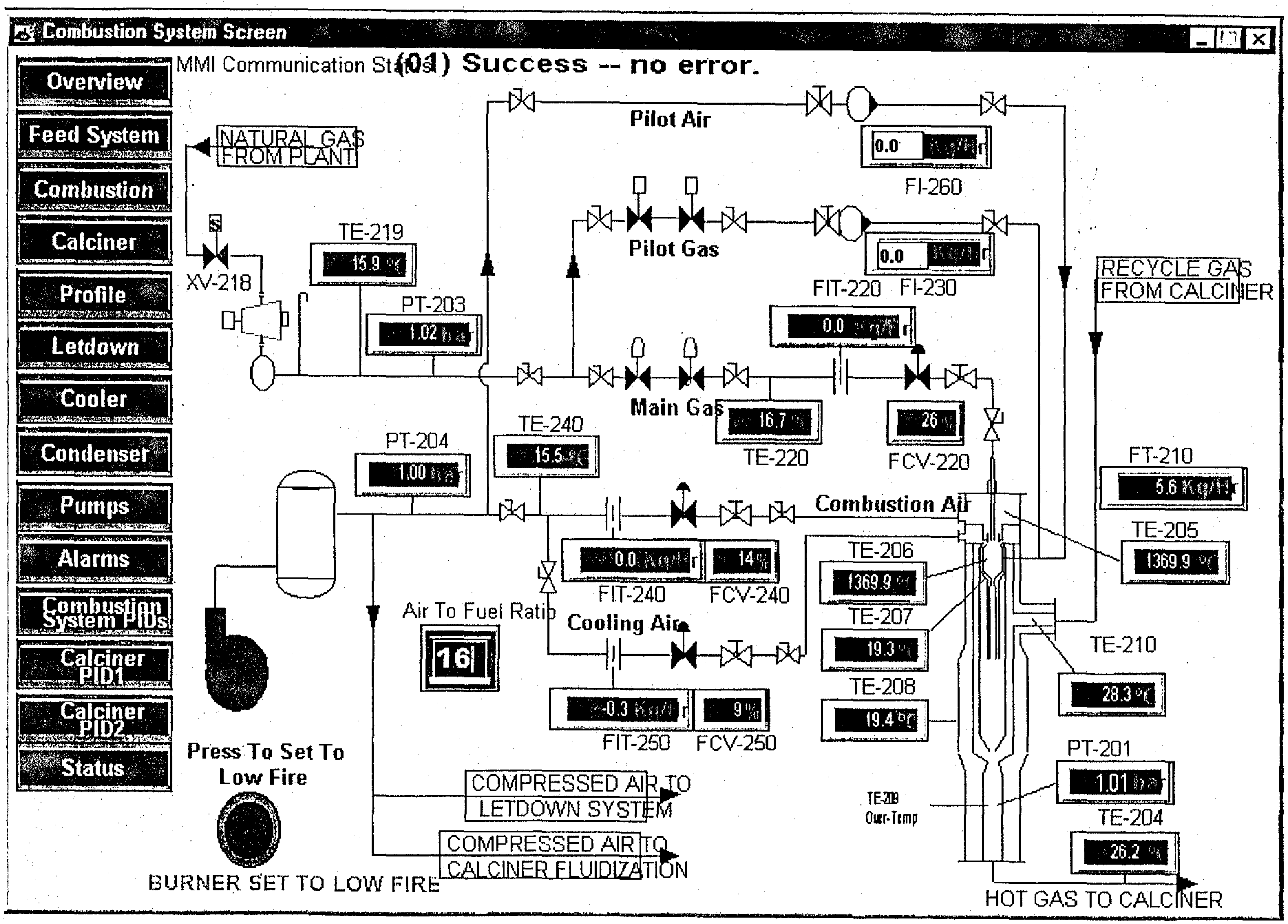
Location — Ground Level

APPENDIX E

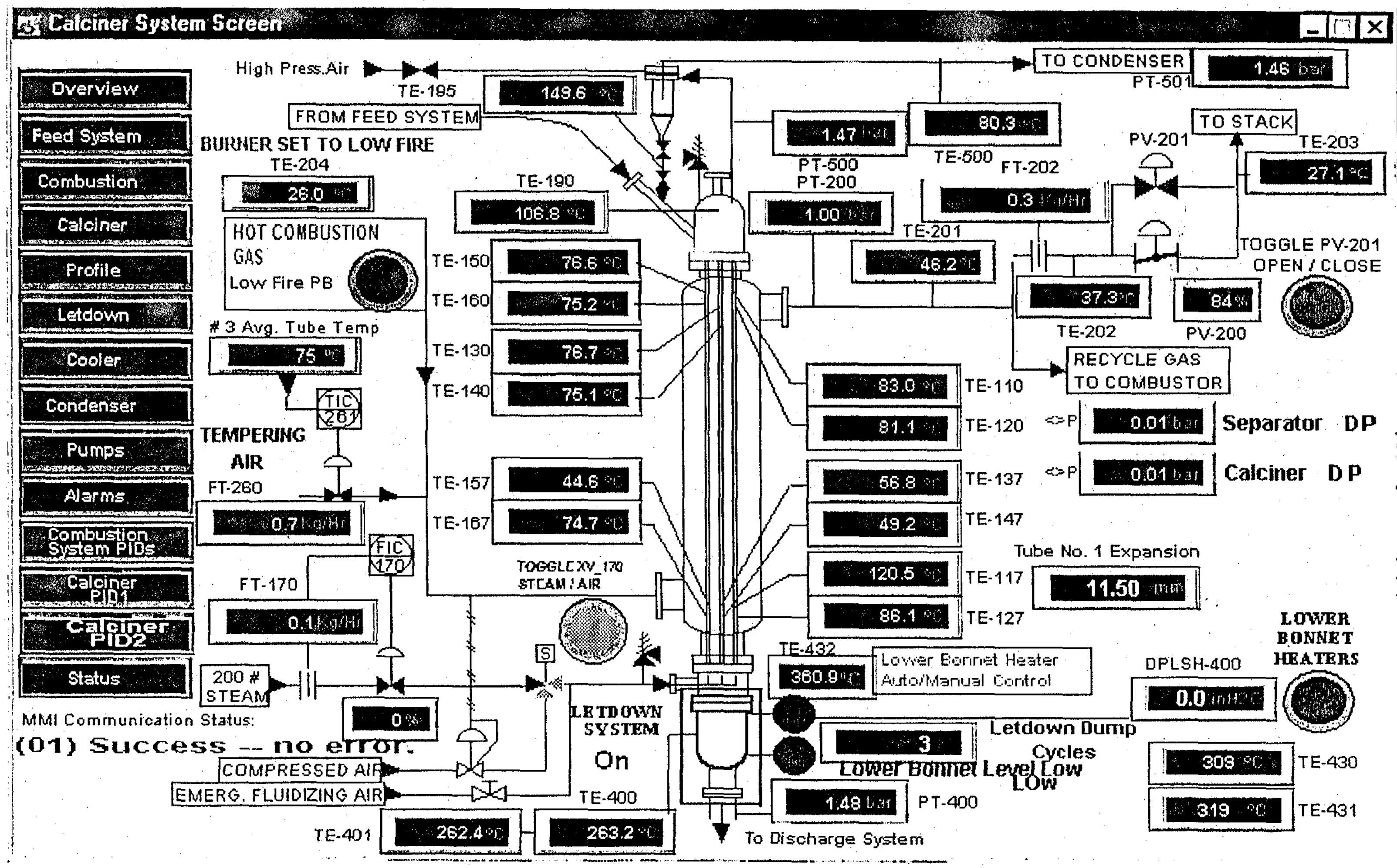
Allen Bradley ControlView Screens

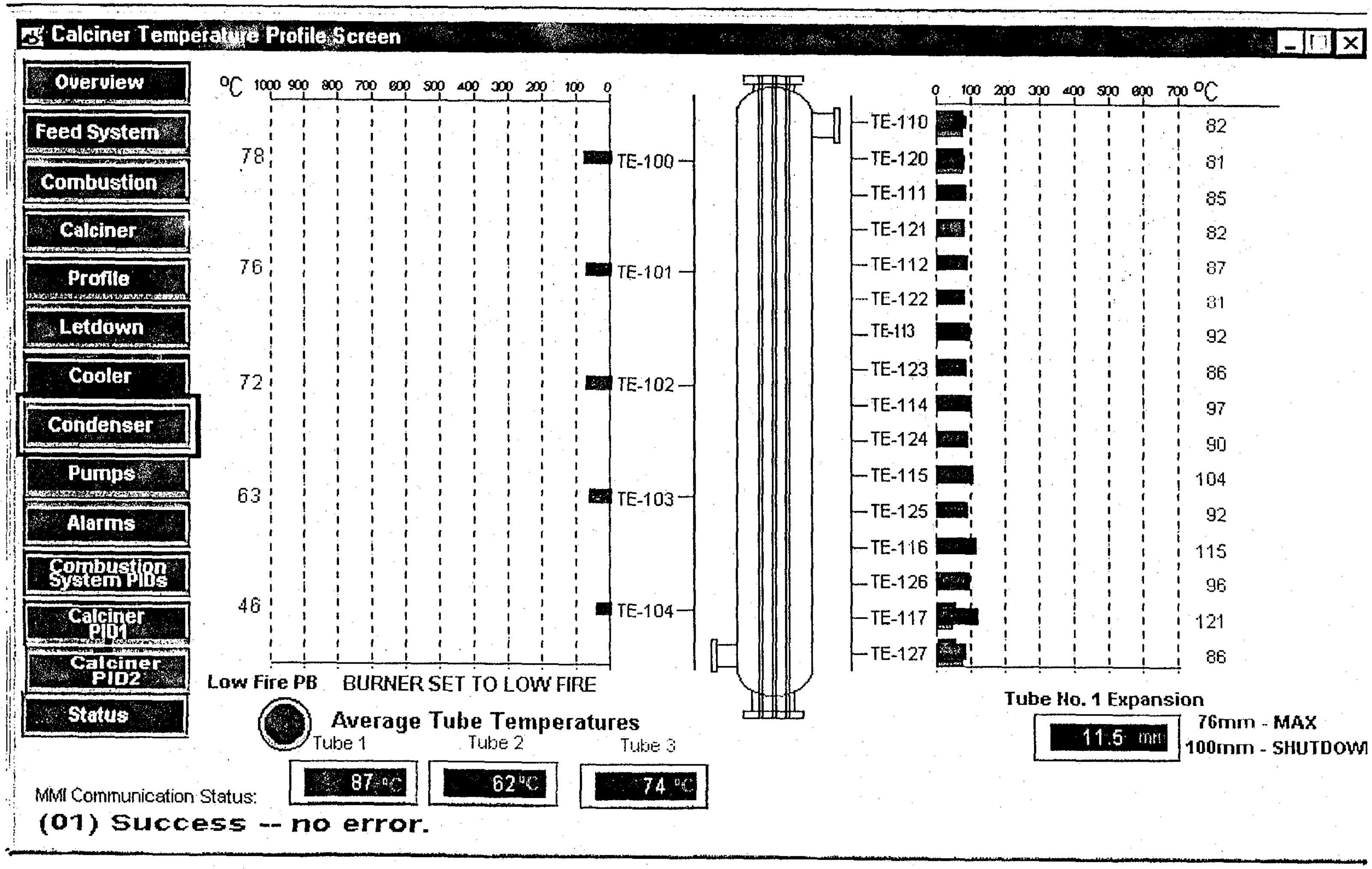

Calciner System Overview Screen
Overview
Feed System
Combustion
Calciner
Profile
Letdown
Cooler
Condenser
Pumps
Alarms
Combustion System PIDs
Calciner PID1
Calciner PID2
Status
MMI Communication
(01) Success -- no error.


Calciner Feed System Status Screen
Overview
Feed System
Combustion
Calciner
Profile
Letdown
Cooler
Condenser
Pumps
Alarms
Combustion System PIDs
Calciner PID1
Calciner PID2
Status
MMI Communication Status:
(01) Success -- no error.
I.A
XV-301
Press To Toggler On/Off Heated Screw Steam
STEAM IN
ST-300
SIC 300
Press To Toggle Screw Feeder Drive Enable
TE-302
26.6
Dome Valve No. 1
Vent Valve
546.1
Macauber Dump Cycles
0
Balance Valve
Press To Toggle Controlveyor System to Auto
Press To Toggle Air / Steam Convey Solenoid Control
Macauber Drain Inject. Vessel
Dome Valve No. 2
TE-303
70.0
Totalized ST-301T
188.7
ST-301
0.0
SIC 301
FIC 311
FT-310
-0.4
PT-306
1.04
TE-304
102.1
200# STEAM
XV-316
Air
XV-315
100
Air/Steam Convey Valve
BURNER SET TO LOW FIRE
HYDRATE TO CALC

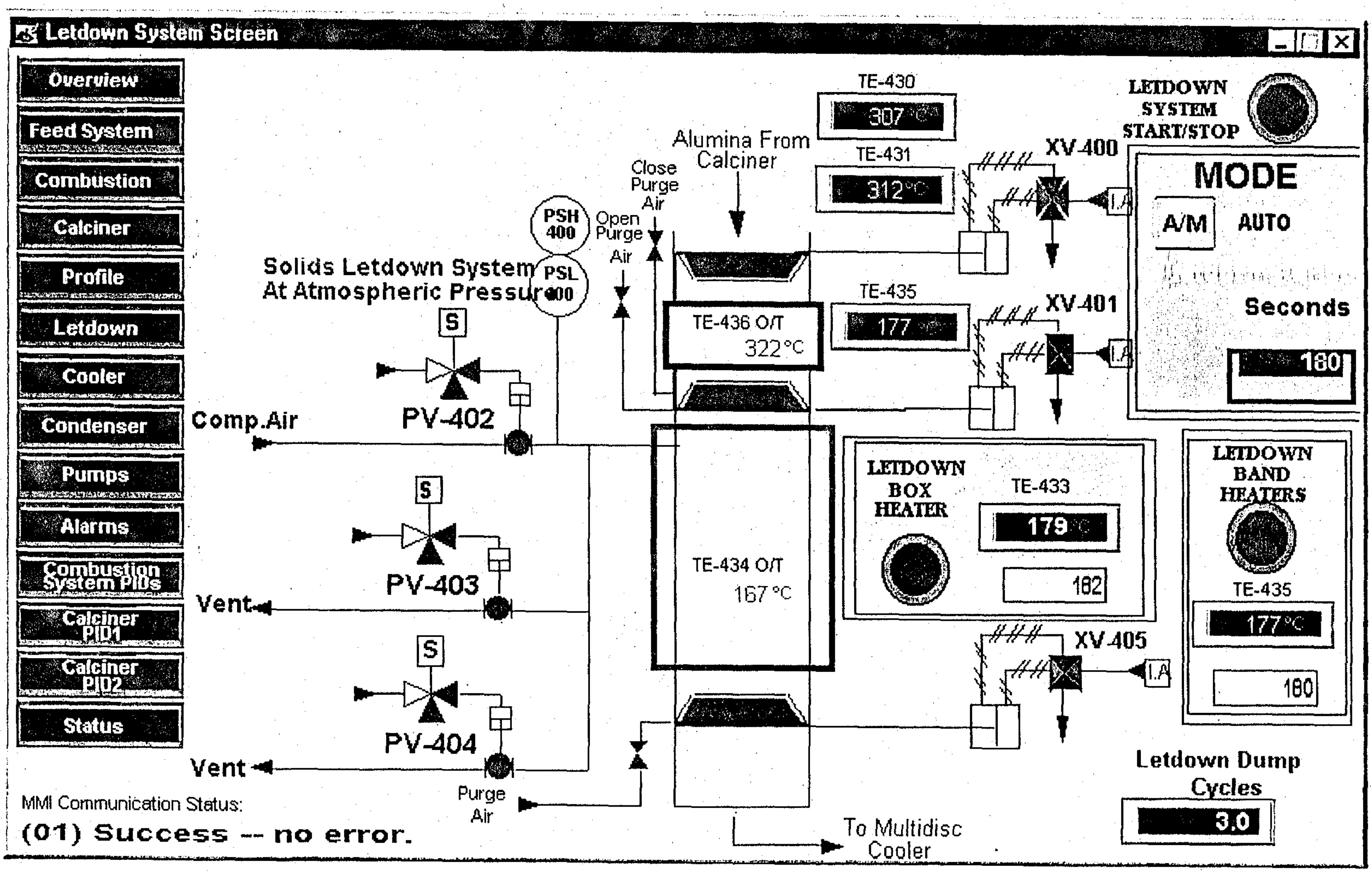


Combustion System Screen
MMI Communication Status (01) Success -- no error.
Overview
Feed System
Combustion
Calciner
Profile
Letdown
Cooler
Condenser
Pumps
Alarms
Combustion System PIDs
Calciner PID1
Calciner PID2
Status
NATURAL GAS FROM PLANT
XV-218
TE-219
15.9
PT-203
1.02
PT-204
1.00
TE-240
15.5
Pilot Air
Pilot Gas
Main Gas
FI-260
0.0
FI-230
0.0
FIT-220
0.0
TE-220
16.7
FCV-220
26
Combustion Air
FIT-240
0.0
FCV-240
14
Cooling Air
FIT-250
-0.3
FCV-250
9
Air To Fuel Ratio
16
RECYCLE GAS FROM CALCINER
FT-210
5.6
TE-205
1369.9
TE-206
1369.9
TE-207
19.3
TE-208
19.4
TE-210
28.3
PT-201
1.01
TE-204
26.2
Press To Set To Low Fire
BURNER SET TO LOW FIRE
COMPRESSED AIR TO LETDOWN SYSTEM
COMPRESSED AIR TO CALCINER FLUIDIZATION
HOT GAS TO CALCINER

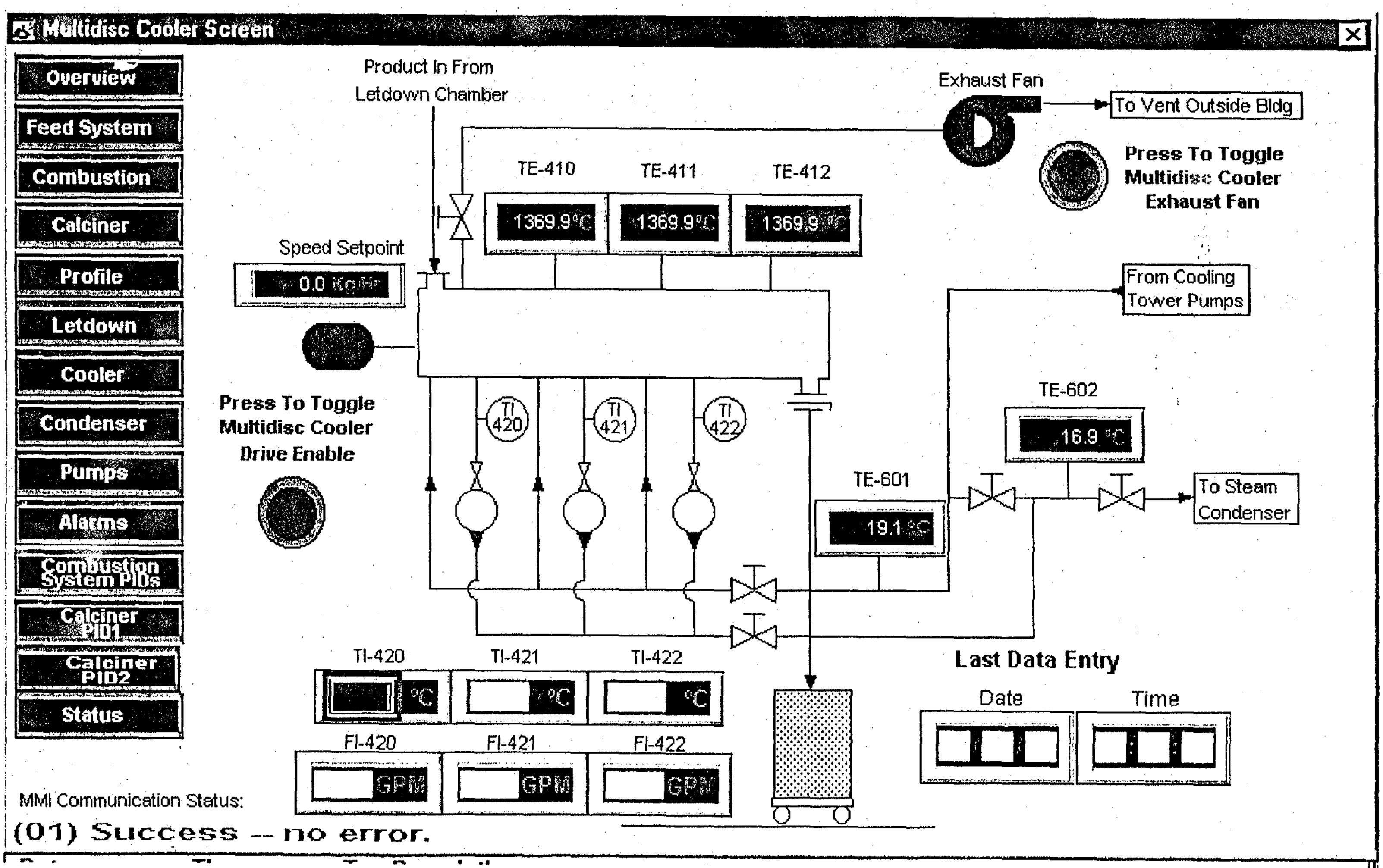

Calciner System Screen
Overview
Feed System
Combustion
Calciner
Profile
Letdown
Cooler
Condenser
Pumps
Alarms
Combustion System PIDs
Calciner PID1
Calciner PID2
Status
MMI Communication Status:
(01) Success -- no error.
High Press.Air
TE-195
149.6 °C
FROM FEED SYSTEM
BURNER SET TO LOW FIRE
TE-204
26.0 °C
HOT COMBUSTION GAS
Low Fire PB
3 Avg. Tube Temp
75 °C
TIC 261
TEMPERING AIR
FT-260
0.7 kg/Hr
FIC 170
FT-170
0.1 kg/Hr
200 # STEAM
0 %
COMPRESSED AIR
EMERG. FLUIDIZING AIR
TE-401
262.4 °C
TE-400
263.2 °C
TE-190
106.8 °C
TE-150
76.6 °C
TE-160
75.2 °C
TE-130
76.7 °C
TE-140
75.1 °C
TE-157
44.6 °C
TE-167
74.7 °C
TOGGLE XV_170 STEAM / AIR
LETDOWN SYSTEM
On
TO CONDENSER
PT-501
1.46 bar
1.47 bar
PT-500
PT-200
1.00 bar
80.3 °C
TE-500
FT-202
0.3 kg/Hr
TE-201
46.2 °C
PV-201
TO STACK
TE-203
27.1 °C
TOGGLE PV-201 OPEN / CLOSE
37.3 °C
TE-202
84 %
PV-200
RECYCLE GAS TO COMBUSTOR
83.0 °C
TE-110
81.1 °C
TE-120
<>P
0.01 bar
Separator DP
<>P
0.01 bar
Calciner DP
56.8 °C
TE-137
49.2 °C
TE-147
Tube No. 1 Expansion
11.50 mm
120.5 °C
TE-117
86.1 °C
TE-127
TE-432
360.9 °C
Lower Bonnet Heater Auto/Manual Control
DPLSH-400
0.0 inH2O
LOWER BONNET HEATERS
3
Letdown Dump Cycles
Lower Bonnet Level Low
LOW
1.48 bar
PT-400
To Discharge System
309 °C
TE-430
319 °C
TE-431

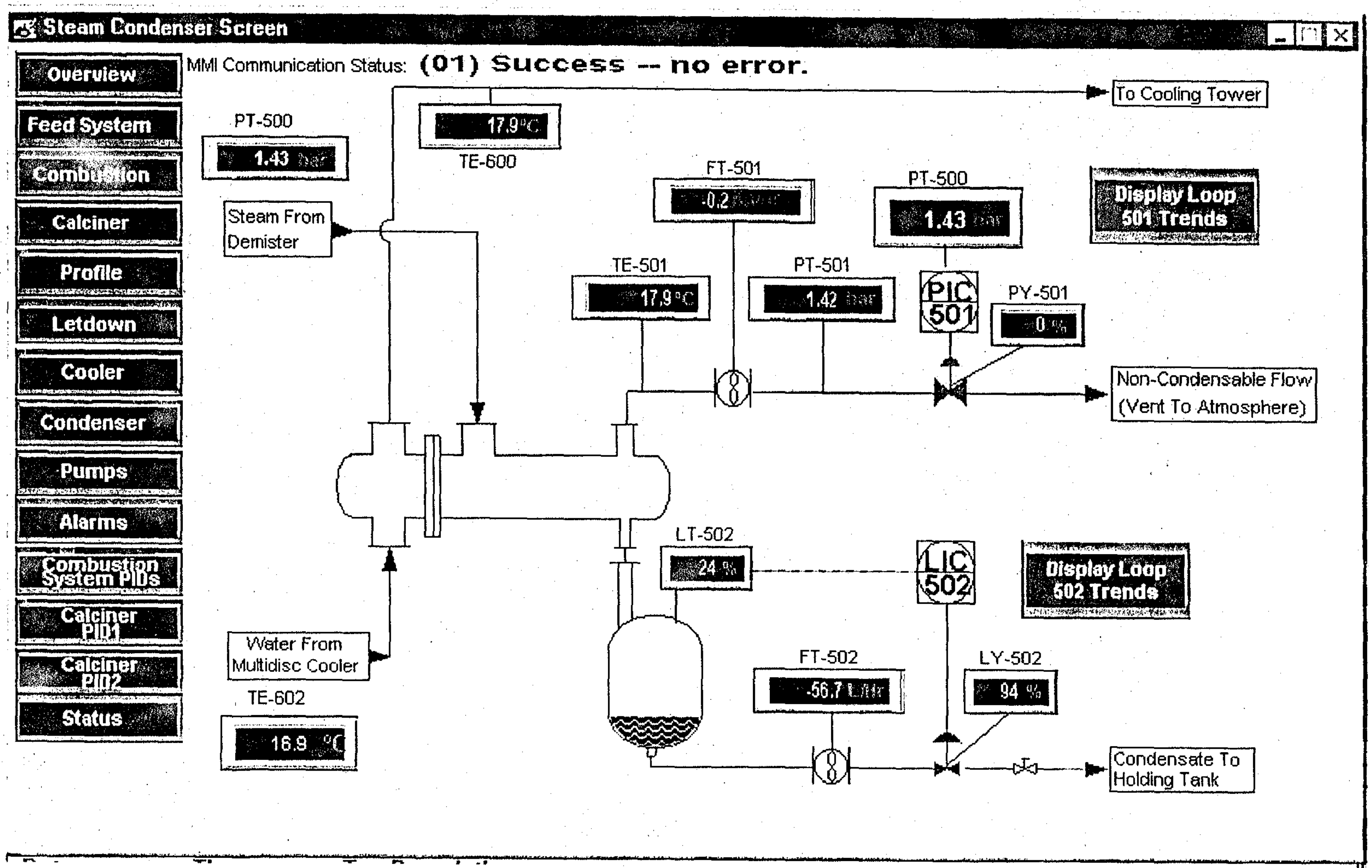


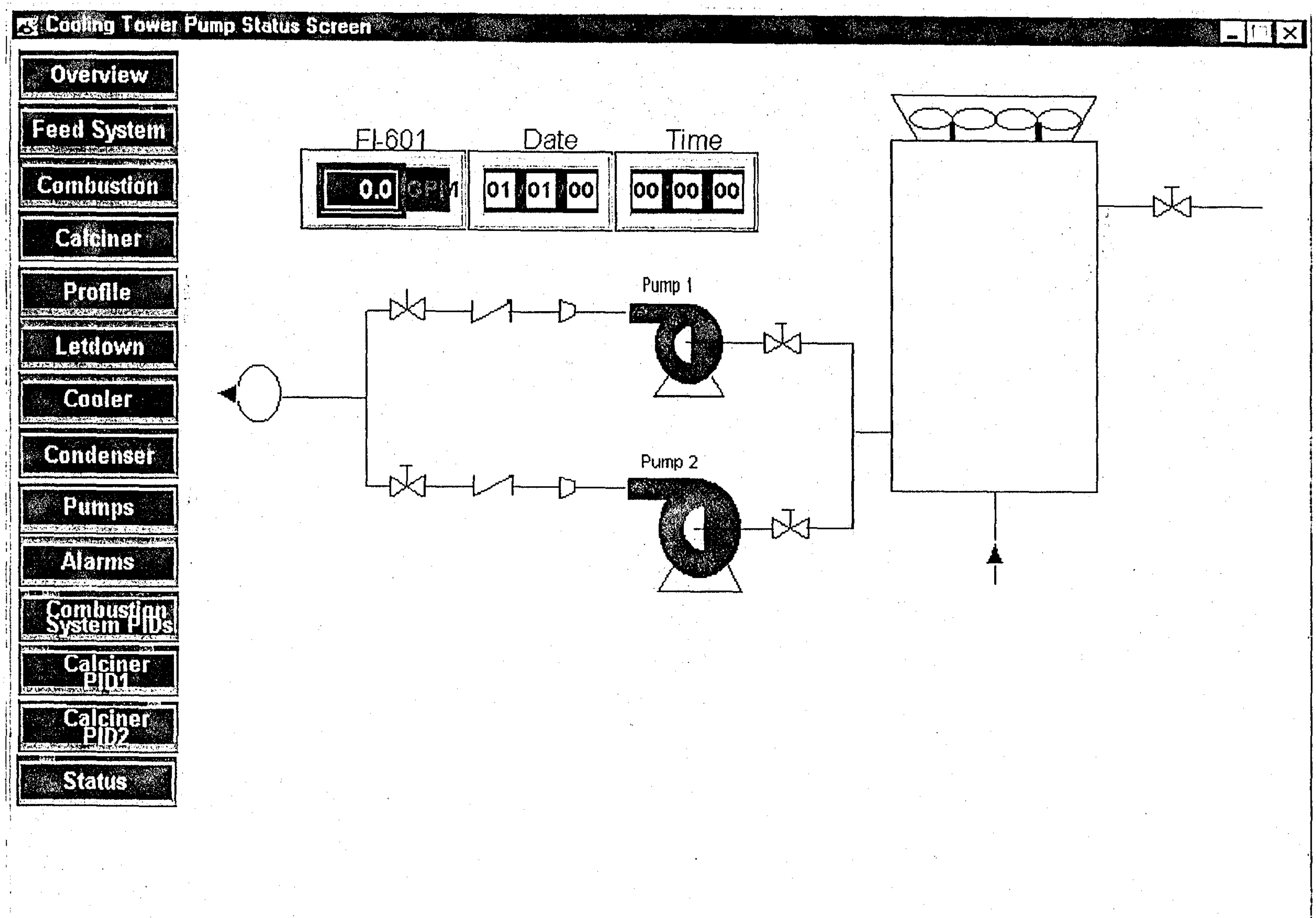
Calciner Temperature Profile Screen
Overview
Feed System
Combustion
Calciner
Profile
Letdown
Cooler
Condenser
Pumps
Alarms
Combustion System PIDs
Calciner PID1
Calciner PID2
Status
°C 1000 900 800 700 600 500 400 300 200 100 0
78 TE-100
76 TE-101
72 TE-102
63 TE-103
46 TE-104
Low Fire PB
BURNER SET TO LOW FIRE
Average Tube Temperatures
Tube 1 87 °C
Tube 2 62 °C
Tube 3 74 °C
0 100 200 300 400 500 600 700 °C
TE-110 82
TE-120 81
TE-111 85
TE-121 82
TE-112 87
TE-122 81
TE-113 92
TE-123 86
TE-114 97
TE-124 90
TE-115 104
TE-125 92
TE-116 115
TE-126 96
TE-117 121
TE-127 86
Tube No. 1 Expansion
11.5 mm
76mm - MAX
100mm - SHUTDOWI
MMI Communication Status:
(01) Success -- no error.

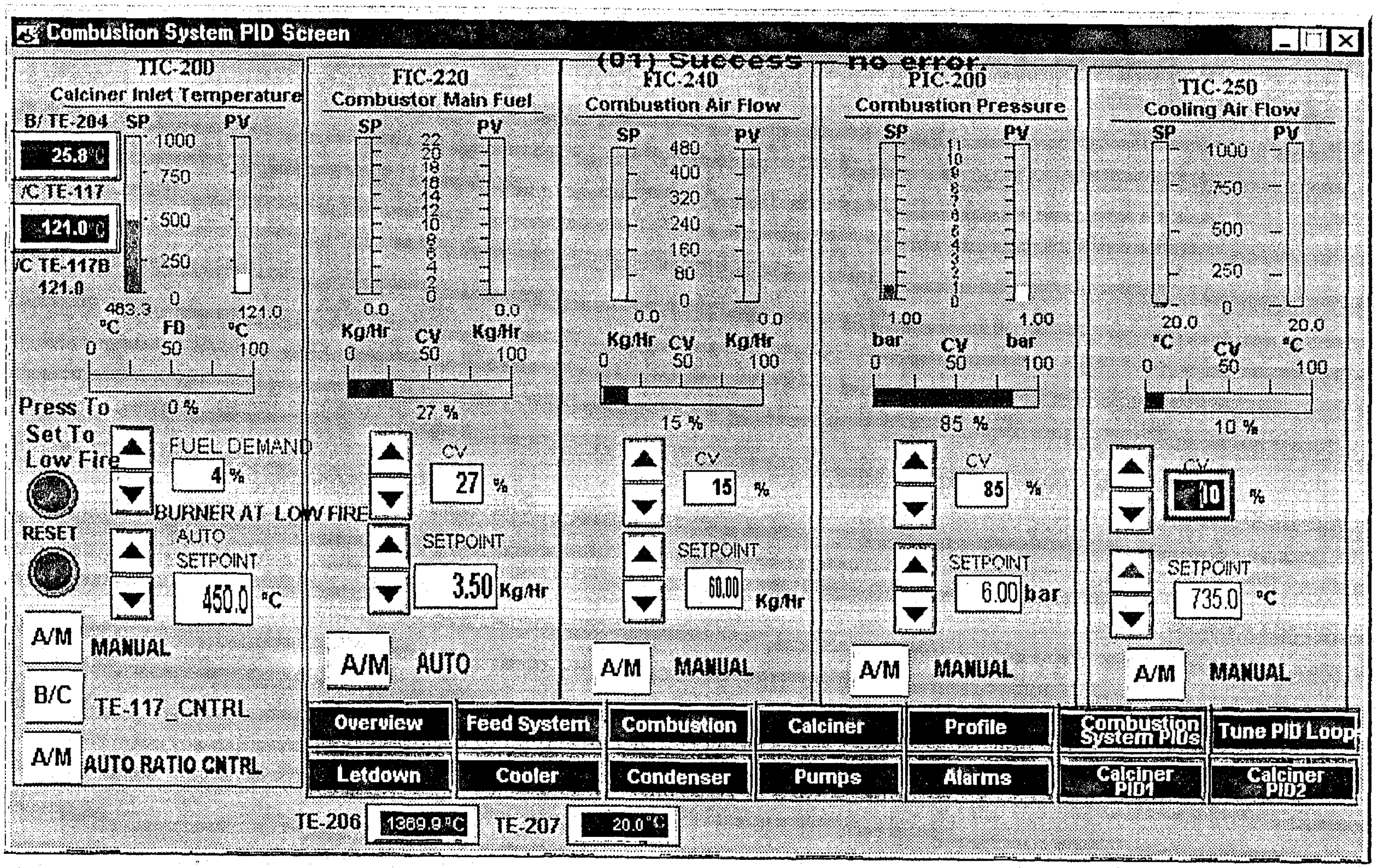


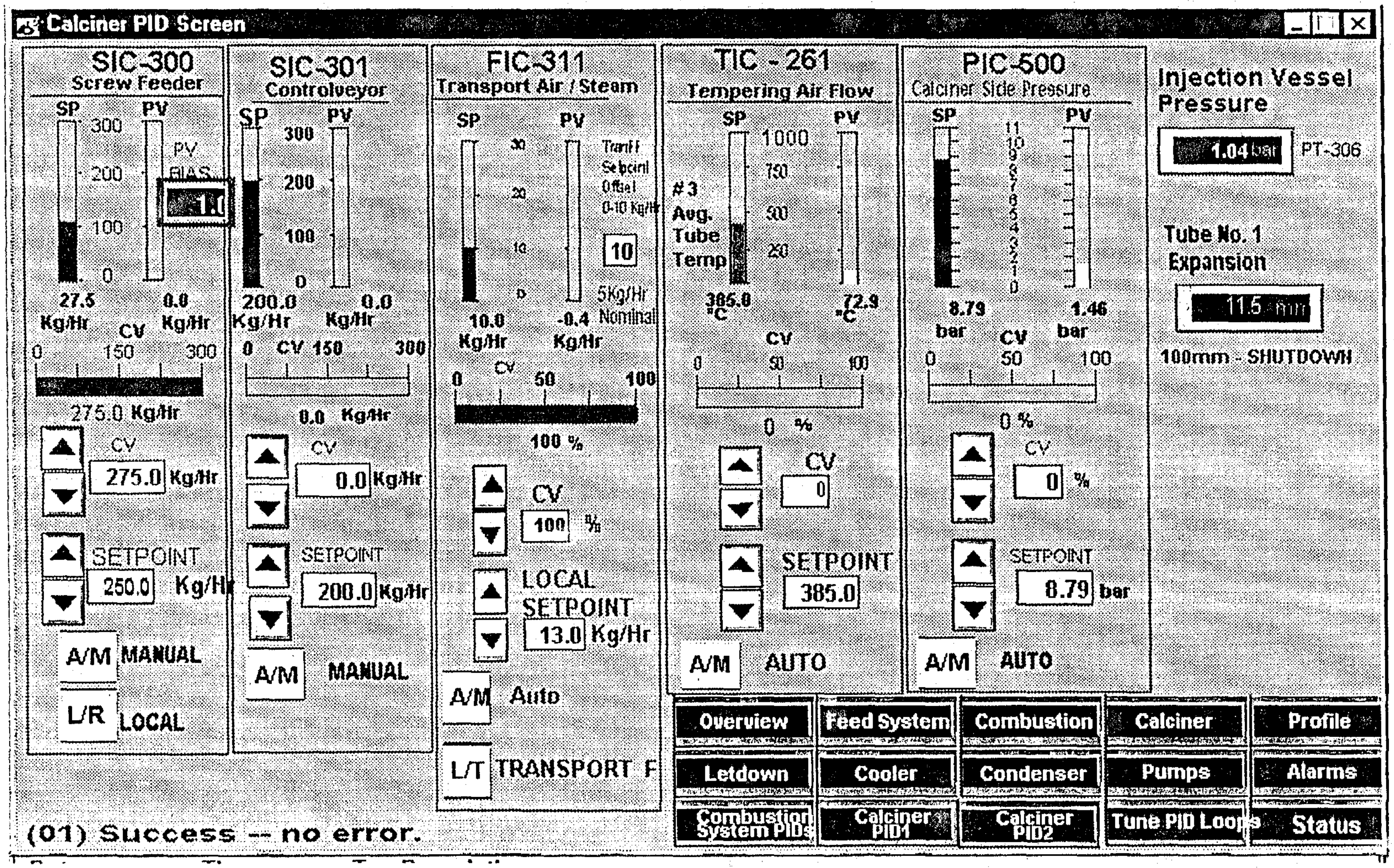
Letdown System Screen
Overview
Feed System
Combustion
Calciner
Profile
Letdown
Cooler
Condenser
Pumps
Alarms
Combustion System PIDs
Calciner PID1
Calciner PID2
Status
Solids Letdown System At Atmospheric Pressure
PSH 400
PSL 400
Comp.Air
PV-402
PV-403
PV-404
Vent
Vent
Purge Air
Open Purge Air
Close Purge Air
Alumina From Calciner
TE-436 O/T 322 °C
TE-434 O/T 167 °C
To Multidisc Cooler
TE-430 307 °C
TE-431 312 °C
TE-435 177
XV-400
XV-401
XV-405
I.A
LETDOWN BOX HEATER
TE-433 179 °C
182
LETDOWN SYSTEM START/STOP
MODE
A/M
AUTO
Seconds
180
LETDOWN BAND HEATERS
TE-435 177 °C
180
Letdown Dump Cycles
3.0
MMI Communication Status:
(01) Success -- no error.

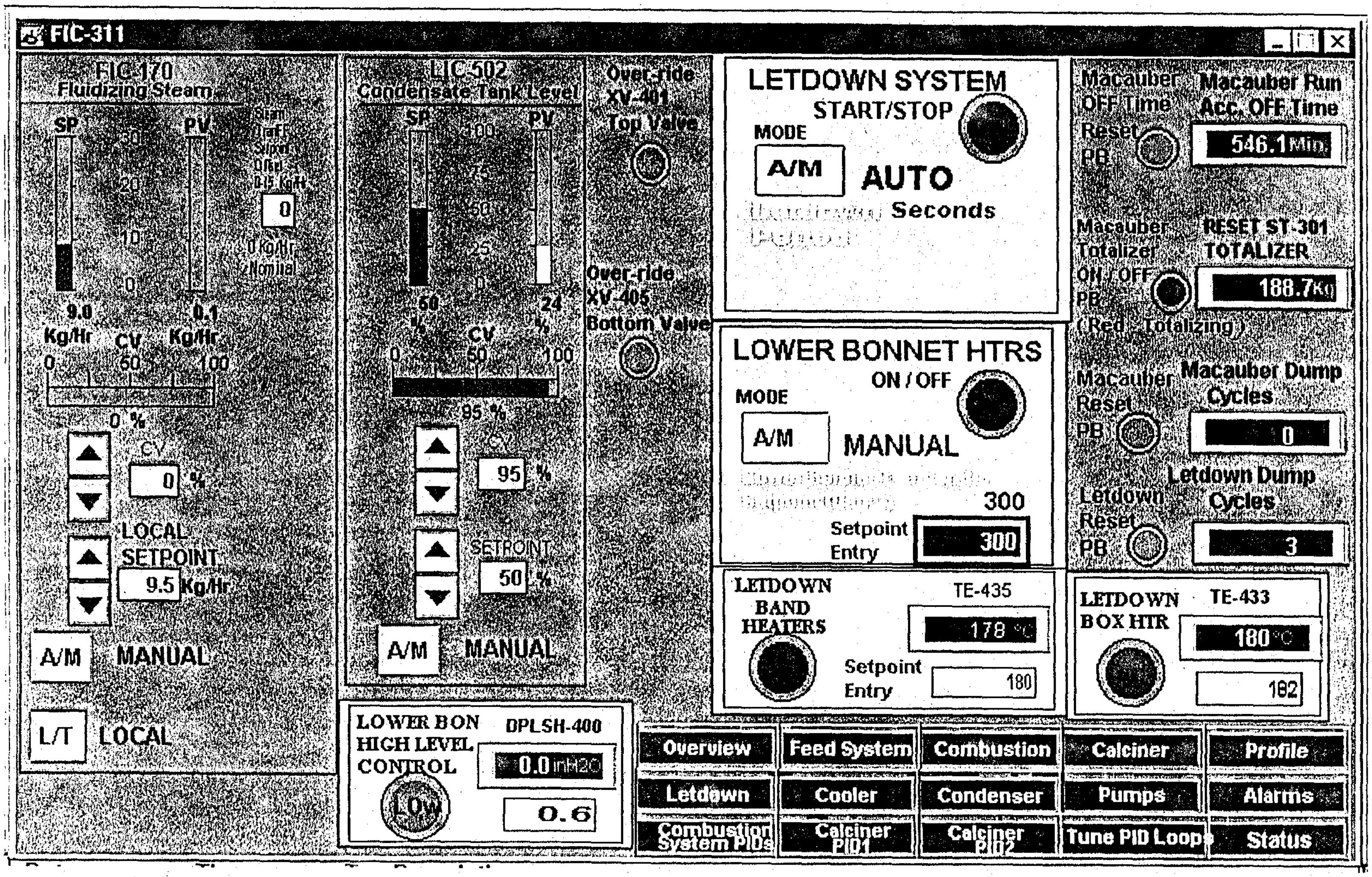

Multidisc Cooler Screen
Overview
Feed System
Combustion
Calciner
Profile
Letdown
Cooler
Condenser
Pumps
Alarms
Combustion System PIDs
Calciner PID1
Calciner PID2
Status
Product In From Letdown Chamber
Exhaust Fan
To Vent Outside Bldg
Press To Toggle Multidisc Cooler Exhaust Fan
TE-410
TE-411
TE-412
1369.9°C
1369.9°C
Speed Setpoint
0.0
From Cooling Tower Pumps
Press To Toggle Multidisc Cooler Drive Enable
TI 420
TI 421
TI 422
TE-602
16.9
TE-601
19.1
To Steam Condenser
TI-420
TI-421
TI-422
°C
Last Data Entry
Date
Time
FI-420
FI-421
FI-422
GPM
MMI Communication Status:
(01) Success -- no error.


Steam Condenser Screen
Overview
Feed System
Combustion
Calciner
Profile
Letdown
Cooler
Condenser
Pumps
Alarms
Combustion System PIDs
Calciner PID1
Calciner PID2
Status
MMI Communication Status: (01) Success -- no error.
To Cooling Tower
PT-500
1.43 bar
17.9°C
TE-600
Steam From Demister
FT-501
-0.2
PT-500
1.43 bar
Display Loop 501 Trends
TE-501
17.9°C
PT-501
1.42 bar
PIC 501
PY-501
0 %
Non-Condensable Flow (Vent To Atmosphere)
LT-502
24 %
LIC 502
Display Loop 502 Trends
Water From Multidisc Cooler
TE-602
16.9 °C
FT-502
-56.7
LY-502
94 %
Condensate To Holding Tank


Cooling Tower Pump Status Screen
Overview
Feed System
Combustion
Calciner
Profile
Letdown
Cooler
Condenser
Pumps
Alarms
Combustion System PIDs
Calciner PID1
Calciner PID2
Status
FI-601
0.0 GPM
Date
01 01 00
Time
00 00 00
Pump 1
Pump 2


Combustion System PID Screen
(01) Success -- no error.
TIC-200
Calciner Inlet Temperature
B/ TE-204
25.8 °C
/C TE-117
121.0 °C
/C TE-117B
121.0
SP PV
1000
750
500
250
0
483.3
121.0
°C FD °C
0 50 100
0 %
Press To Set To Low Fire
RESET
FUEL DEMAND
4 %
BURNER AT LOW FIRE
AUTO SETPOINT
450.0 °C
A/M MANUAL
B/C TE-117_CNTRL
A/M AUTO RATIO CNTRL
FIC-220
Combustor Main Fuel
SP PV
0.0 Kg/Hr CV 0.0 Kg/Hr
0 50 100
27 %
CV 27 %
SETPOINT 3.50 Kg/Hr
A/M AUTO
FIC-240
Combustion Air Flow
SP PV
480
400
320
240
160
80
0
0.0 Kg/Hr CV 0.0 Kg/Hr
0 50 100
15 %
CV 15 %
SETPOINT 60.00 Kg/Hr
A/M MANUAL
PIC-200
Combustion Pressure
SP PV
1.00 bar CV 1.00 bar
0 50 100
85 %
CV 85 %
SETPOINT 6.00 bar
A/M MANUAL
TIC-250
Cooling Air Flow
SP PV
1000
750
500
250
0
20.0 °C CV 20.0 °C
0 50 100
10 %
CV 10 %
SETPOINT 735.0 °C
A/M MANUAL
Overview
Feed System
Combustion
Calciner
Profile
Combustion System PIDs
Tune PID Loop
Letdown
Cooler
Condenser
Pumps
Alarms
Calciner PID1
Calciner PID2
TE-206 1369.9 °C
TE-207 20.0 °C


Calciner PID Screen
SIC-300
Screw Feeder
SP
PV
PV BIAS
27.5 Kg/Hr
0.0 Kg/Hr
CV
275.0 Kg/Hr
SETPOINT
250.0
Kg/Hr
A/M MANUAL
L/R LOCAL
SIC-301
Controlveyor
200.0 Kg/Hr
0.0 Kg/Hr
0.0 Kg/Hr
200.0 Kg/Hr
A/M MANUAL
FIC-311
Transport Air / Steam
Tranf Setpoint Offset 0-10 Kg/Hr
10
5Kg/Hr Nominal
10.0 Kg/Hr
-0.4 Kg/Hr
100 %
CV
100 %
LOCAL SETPOINT
13.0 Kg/Hr
A/M Auto
L/T TRANSPORT F
TIC - 261
Tempering Air Flow
#3 Avg. Tube Temp
385.0 °C
72.9 °C
0 %
SETPOINT
385.0
A/M AUTO
PIC-500
Calciner Side Pressure
8.79 bar
1.46 bar
0 %
SETPOINT
8.79 bar
A/M AUTO
Injection Vessel Pressure
1.04 bar
PT-306
Tube No. 1 Expansion
11.5 mm
100mm - SHUTDOWN
Overview
Feed System
Combustion
Calciner
Profile
Letdown
Cooler
Condenser
Pumps
Alarms
Combustion System PIDs
Calciner PID1
Calciner PID2
Tune PID Loops
Status
(01) Success -- no error.


FIC-311
FIC-170
Fluidizing Steam
SP
PV
9.0 Kg/Hr
0.1 Kg/Hr
CV
0 %
0
LOCAL SETPOINT
9.5 Kg/Hr
A/M
MANUAL
L/T
LOCAL
LIC-502
Condensate Tank Level
SP
PV
CV
95 %
95
SETPOINT
50
A/M
MANUAL
Over-ride XV-401 Top Valve
Over-ride XV-405 Bottom Valve
LETDOWN SYSTEM
START/STOP
MODE
A/M
AUTO
Seconds
LOWER BONNET HTRS
ON / OFF
MODE
A/M
MANUAL
300
Setpoint Entry
300
LETDOWN BAND HEATERS
TE-435
178 °C
Setpoint Entry
180
LETDOWN BOX HTR
TE-433
180 °C
182
Macauber OFF Time Reset PB
Macauber Run Acc. OFF Time
546.1 Min
Macauber Totalizer ON / OFF PB (Red - Totalizing)
RESET ST-301 TOTALIZER
188.7 Kg
Macauber Reset PB
Macauber Dump Cycles
0
Letdown Reset PB
Letdown Dump Cycles
3
LOWER BON HIGH LEVEL CONTROL
DPLSH-400
0.0 inH2O
Low
0.6
Overview
Feed System
Combustion
Calciner
Profile
Letdown
Cooler
Condenser
Pumps
Alarms
Combustion System PIDs
Calciner PID1
Calciner PID2
Tune PID Loops
Status